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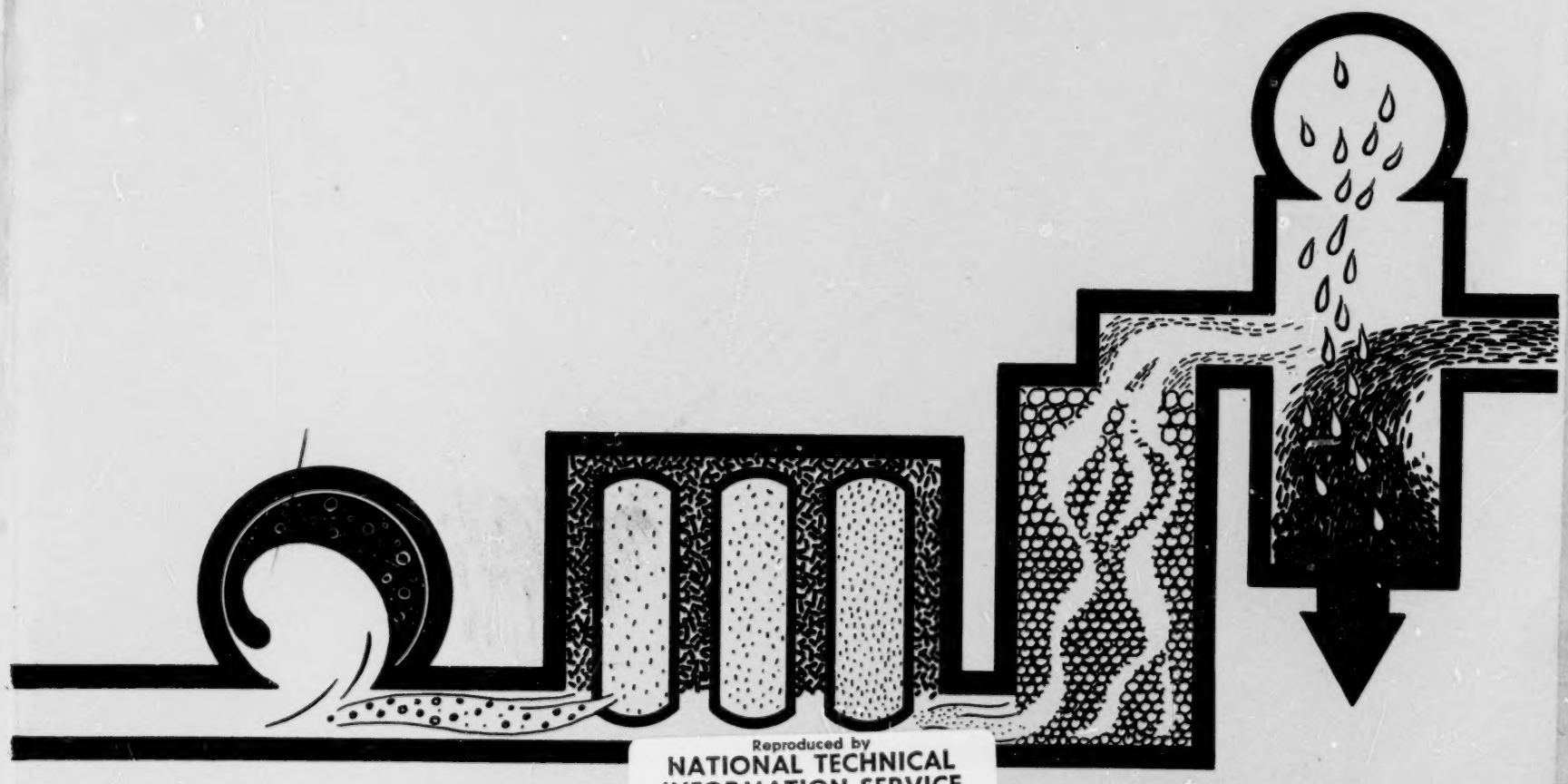
FEASIBILITY OF TREATING WASTEWATER BY DISTILLATION



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16. Abstract A long tube, vertical, three-effect evaporator was subjected to tests to determine effects of feedwater quality and evaporation conditions on product water quality, post evaporation polishing, and evaporator tube scaling. Feedwaters tested included effluents from extended aeration, trickling filter, and contact stabilization units. All units treated the same influent wastes. An odor free product could not be produced at temperatures between 112°F and 290°F, with the odor increasing as either the evaporator temperature or the feedwater COD increased. Preliminary studies demonstrated that while aeration could not always remove all the odor, activated carbon adsorption could. pH adjustment was shown to be incapable of controlling ammonia, but ion exchange provided sufficient removal. Scaling problems were prohibitive with trickling filter effluent but nearly negligible with extended aeration effluent. Increased efficiencies, at higher operating temperatures and negligible boiling point rise indicated that wastewater evaporation should be more economical than seawater evaporation. (Lowry-Texas)

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FEASIBILITY OF TREATING WASTEWATER BY DISTILLATION

by

University of Florida
Gainesville, Florida 32601

for the

ENVIRONMENTAL PROTECTION AGENCY

FWPCA

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ABSTRACT

The technical feasibility of evaporation of municipal sewage treatment plant effluent for the purpose of water reuse was investigated. The equipment used was a long tube vertical (LTV) evaporator. The objectives of the research were to determine the effects of feedwater quality, and evaporation conditions on product water quality, post evaporation polishing, and evaporator tube scaling.

The experimental equipment consisted of a three-effect evaporator constructed of 316 stainless steel. Each effect contained a single one-inch-diameter tube with a 14-ft. effective heated length. The evaporator could be operated as a single-effect or triple-effect unit. Possible operating conditions varied from around 28 in. Hg vacuum to about 55 psig. All evaporator feedwater was first acidified and degassed under vacuum to remove dissolved gases.

Feedwaters tested in the evaporator included extended aeration effluent, high rate trickling filter effluent, and contact stabilization effluent. All three units treated raw waste from a large university complex.

Results showed that an odor free product could not be produced from any of the three feedwaters under any operating condition from 112°F to 290°F. The odor intensity increased as the evaporator operating temperature and/or the chemical oxygen demand of the feedwater increased. In all cases, post treatment with activated carbon removed all odors. Aeration would not remove all odors. Product produced under pressure conditions using trickling filter feed contained significantly more organic contamination than any other products. Product contamination by ammonia could not be controlled by adjusting the pH of the feed in the range 5.1 to 8.7. However, ammonia in the product water was removed by ion exchange.

The scaling evaluations were carried out under pressure conditions using extended aeration and trickling filter effluent. Trickling filter feedwater was judged unsuitable because of excessive scaling. Scaling problems with extended aeration feedwater were minor, and with post treatment by activated carbon, a high quality product water was produced. Trickling filter feedwater gave more severe scaling problems, mostly because of the use of more sulfuric acid in the degassing pretreatment.

Because of increased efficiencies due to higher operating temperatures and negligible boiling point rise, wastewater evaporation should be more economical than sea water evaporation.

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CONTENTS

	Page
Conclusions	1
Recommendations	3
Introduction	5
Theoretical Background	9
Research Plan	21
Description of Equipment	27
Experimental Results	37
Discussion of Results	69
Appendices	75
References	81

FIGURES

NO.		PAGE
1	DISTRIBUTION OF NH_3 AND NH_4^+ AS A FUNCTION OF pH AND TEMPERATURE	12
2	CALCIUM SULFATE SOLUBILITY	16
3	OVERALL FLOWSHEET	28
4	DEGASSING SYSTEM FLOWSHEET	29
5	EVAPORATOR FLOWSHEET	31
6	EVAPORATOR	32
7	VAPOR-LIQUID SEPARATOR	33
8	PRODUCT RECEIVER SYSTEM FLOWSHEET	34
9	CONDUCTIVITY VS. AMMONIA CONCENTRATION	52
10	PRODUCT RATES, FIRST EXTENDED AERATION EFFLUENT RUN	55
11	OVERALL HEAT TRANSFER COEFFICIENTS	60

TABLES

No.		Page
1	Dissociation Constants for Aqueous Ammonia and Water	11
2	Analytical Tests on Liquid Samples	24
3	Analytical Tests on Scale	26
4	Evaporator Operating Conditions for Initial Single Stage Tests	39
5	Initial Tests, Extended Aeration Effluent, Vacuum Conditions	40
6	Initial Tests, Extended Aeration Effluent, Atmospheric Conditions	41
7	Initial Tests, Extended Aeration Effluent, Pressure Conditions	42
8	Initial Tests, Contact Stabilization Effluent, Vacuum Conditions	44
9	Initial Tests, Contact Stabilization Effluent, Atmospheric Conditions	45
10	Initial Tests, Contact Stabilization Effluent, Pressure Conditions	46
11	Initial Tests, Trickling Filter Effluent, Vacuum Conditions	47
12	Initial Tests, Trickling Filter Effluent, Atmospheric Conditions	48
13	Initial Tests, Trickling Filter Effluent, Pressure Conditions	49
14	Typical Analysis of Tap Water, Gainesville, Florida	53
15	Analytical Results - First Extended Aeration Effluent Run	57
16	Scale from First Extended Aeration Effluent Run	58
17	Analytical Results - Trickling Filter Effluent Run	59
18	Scale from Trickling Filter Effluent Run	61

No.		Page
19	Activated Carbon Treatment of Products from Trickling Filter Effluent Run	61
20	Analytical Results - Second Extended Aeration Effluent Run	63
21	Bacteriological Tests - Second Extended Aeration Effluent Run	64
22	Scale from Second Extended Aeration Effluent Run	65
23	Analytical Results - Three Effect Ammonia Distribution Tests	67
24	Campus Sewage Treatment Plant Test Results, High Rate Trickling Filter	75
25	Campus Sewage Treatment Plant Test Results, Contact Stabilization Plant	76

CONCLUSIONS

Based on the results of this study the following conclusions have been drawn:

1. Product water produced by evaporation of extended aeration, contact stabilization, or high rate trickling filter effluent is not acceptable for reuse without further treatment because of organic odor carry-over from the feed.

2. Product odor tends to increase with increasing evaporator operating temperature in the range 112-288^{o}F.

3. The odor of products produced from secondary effluents can be effectively removed by treatment with activated carbon. However, preliminary indications are that the odors cannot be reliably removed by aeration.

4. At high temperature (280-290^{o}F) significantly more organic carry-over to the product water occurs with trickling filter effluent feed than with extended aeration or contact stabilization effluent feed.

5. Ammonia in the evaporator feedwater can be expected to contaminate the product water.

6. Removal of ammonia from the product water by ion exchange appears practical.

7. Trickling filter effluent appears to be less suitable as evaporator feedwater than extended aeration or contact stabilization effluent because of greater scaling problems.

8. At temperatures up to 280^{o}F, scaling problems with extended aeration effluent appear to be minor.

9. Evaporation of extended aeration effluent followed by activated carbon treatment gives a completely demineralized odor-free product water suitable for high quality reuse.

10. Because of increased efficiencies due to higher operating temperatures and negligible boiling point rise, wastewater evaporation should be more economical than sea water evaporation.

RECOMMENDATIONS

In order to develop the concept of evaporation of sewage treatment plant effluent to the demonstration plant stage, additional work is recommended in the following areas:

1. Additional longer term pilot plant runs, lasting at least two months, should be made to better define any scaling problems.

2. Additional pilot plant runs should be made at temperatures above 280°F to find the maximum allowable operating temperature. This maximum may be controlled either by product quality or scaling problems.

3. Studies should be made to determine the capacity of activated carbon when used to treat evaporator products for odor removal. This would provide the basis for reliable economic estimates of this post treatment process.

4. Corrosion tests should be made to determine the most economical materials of construction for a full scale unit.

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INTRODUCTION

General Background

The need for improved water pollution control combined with an increasing scarcity of natural waters has led to the concept of wastewater renovation for reuse. One treatment method that has potential for producing a high quality water from wastewater is distillation or evaporation.

Extensive interest has developed in the use of distillation for producing usable water from seawater. Development work by the U. S. Department of Interior's Office of Saline Water and by other organizations around the world has reduced the cost of the process to the point at which water can be produced in some locations at a price competitive with or lower than that for water from other sources.

While much of the technology of seawater distillation can probably be applied to wastewater, there are differences in the two feeds which have an important effect on operating conditions and product quality. Municipal wastewater usually has less than 1000 mg/l dissolved solids compared with 35,000 mg/l for seawater. [The work here has been limited to municipal wastewater of essentially domestic origin. This does not preclude the use of distillation on municipal wastes containing industrial wastes or on industrial waste alone. Some of the conclusions drawn from this study may be applicable to a number of industrial wastes.] The corrosiveness and scaling potential (for materials such as calcium sulfate) of wastewater should be less than for seawater. The boiling point elevation for wastewater is essentially zero. These characteristics of wastewater should allow for higher distillation temperature and higher concentration of the brine or blowdown than is practical with sea water. In contrast to the advantages for distillation of wastewater are the disadvantages of soluble and suspended organic materials which may foul heat transfer surfaces and which may contribute taste and odor to the distillate. Ammonia, which is present to some degree in all municipal wastes, has measurable volatility under usual distilling conditions and, therefore, also affects product quality. Its corrosiveness toward copper presents an additional problem. Phosphate is present in significant quantities. Its presence could be an advantage or disadvantage depending upon its interaction with other materials in the water.

There are a number of important questions that remain to be answered before the feasibility of wastewater distillation can be realistically assessed. The present work was undertaken to answer some of these questions.

Previous Work on Evaporation of Treated Sewage

Previous work on evaporation of sewage treatment plant effluent has been limited to several desk top evaluations using presumably reasonable assumptions and four studies that involved actual experimental work.

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In 1960, Hickman[1] compared distillation with other unit processes for the reclamation of wastewaters at the request of the Advanced Waste Treatment Research Program of the U. S. Public Health Service. Hickman concluded that the relatively high cost of distillation (estimated at $1.00/1,000 gal.) should limit it to processing only part of the water in a total recycle scheme. Other less expensive treatment procedures should be used for the remaining parts of the system. Hickman reported the results of one experimental run using treated sewage from a Rochester, N. Y., sewage disposal plant. The specific evaporator used was not described but operating conditions were reported as 120°F (1.69 psia). In the author's opinion, "the physical properties, flavor, and odor of the distillate were good, and that with mild chlorination, or equivalent, the distillate would be potable and acceptable for municipal reuse." Possible contamination of the distillate samples prevented any valid conclusions based on actual analysis.

In 1963 an economic analysis and a pilot plant study of wastewater distillation were made by Neale.[2] He proposed that distillation be used in parallel with some other form of treatment and that the fraction of the total stream subjected to distillation be the minimum required to prevent build up of inorganics in the overall system. Analysis of data from 22 cities indicated that the fraction of the total stream that would have to be distilled ranged from 23.6% to 63.7% and averaged 42.7%. A pilot plant study was carried out in a long tube vertical (LTV) evaporator utilizing a single tube 14 ft long. The primary objective of the tests was to evaluate tube fouling. No provision was made to condense and collect product water. Using primary effluent and no pretreatment, tube fouling was noted after four days. Operation was at 212°F. A second seven-day run at 212°F was made using secondary effluent with no pretreatment. Again tube fouling was noted. The effervescence noted when acid was added to this scale led Neale to conclude that it was largely calcium carbonate. Subsequent runs were made using secondary effluent treated with inhibited hydrochloric acid for pH adjustment. Feed pH was maintained at less than 5.7 most of the time. An eight day run at 212°F and a twelve day run at 234°F were made with no scaling problems. Neale concluded "that waste treatment by distillation will probably be applicable to renovation of wastewater at temperatures equal to or exceeding those used for sea water conversion, and that raw sewage, after removal of most of the suspended matter, may be used for distillation plant feed."

In June, 1963, American Machine and Foundry, Inc., under contract to the Government, undertook research on flash evaporation of treated sewage.[3] The equipment consisted of an eight-stage flash unit. The still body was steel with titanium heat transfer surfaces. Due to operating problems, very limited testing was carried out. Using secondary effluent under relatively high temperature conditions, 250-300°F, severe fouling of the preheater and some fouling of the feed side of the condenser tubes occurred. The scale appeared to be largely organic. The product water had a strong, disagreeable odor. However, treatment with activated carbon removed the odor.[4-6]

In 1965, O'Connor[7] and co-workers at the Robert A. Taft Sanitary Engineering Center in Cincinnati, Ohio, conducted laboratory scale tests on

distillation of municipal effluents. Batch tests were made using primary effluent, trickling filter effluent, and extended aeration effluent. Temperatures used were 158, 212, and 338°F and pH ranged from 3 to 11. At the lower pH values it was hoped that the ammonia would be retained in the concentrated liquid. At the higher pH values, hopefully all the ammonia would distill over in the first small fraction of product. All feed samples were filtered prior to evaporation to remove solid particles larger than about 5 microns. It was found that ammonia was present in the product water for all feeds with pH above about 3.5. With feeds containing around 7-8 mg/1 NH_3-N, ammonia in concentrations above 1.0 mg/1 NH_3-N typically was found in the first 40-50% of the distillate. Even at pH 11, ammonia in concentrations above 1.0 mg/1 NH_3-N was found in at least the first 30% of the distillate. Hence it appeared impractical to try, on a large scale, either to eliminate ammonia from the distillate by lowering the feed pH or to concentrate all the ammonia in the first small fraction of the distillate by raising the feed pH.

Tests with extended aeration plant effluent having ammonia concentrations less than 2.2 mg/1 NH_3-N showed that, at feed pH values under 5.5, all product fractions had ammonia concentrations less than 1.0 mg/ NH_3-N. No carry-over of nitrate was observed.

Odors were present in all products and seemed to be more "prominent and enduring" at high temperatures. Product odor seemed to be independent of feed pH. Powdered carbon at 1 gm/1 removed all odors from products produced under vacuum and atmospheric conditions. However, carbon treatment only partially removed the odors from the products made at high temperature and pressure conditions.

Economic analyses of distillation of treated wastewaters have been made by several people. Gerster[8] made a desk top study of multistage flash, multiple-effect, and recompression-flash evaporation with respect to how these processes might apply to wastewaters. He concluded that "For all types of equipment the cost for the distillation step alone is somewhat less than for sea water, but inclusion of costs for feed pretreatment and ultimate disposal of blowdown, bring the cost up to about that for seawater."

Stephan[9,10] in 1965 predicted that wastewater evaporation should cost about the same as seawater evaporation. Also, he concluded that a parallel renovation system utilizing evaporation on half of the recycle stream would cost about the same as using electrodialysis on the entire stream.

Objectives of this Research

Review of the previous work on evaporation of sewage treatment plant effluent shows that additional information is needed in several areas. First, very little work has been done on a continuous basis utilizing equipment simulating full scale operations. Also, renovation by evaporation should be considered in relationship to any additional pretreatment

or post treatment processes that may be required to render the water acceptable for reuse.

Based on these needs, the following objectives were set for this research.

1. To define the relationships between product water quality, feedwater quality, and evaporation conditions.

2. To define the relationships between post evaporation polishing required, feedwater quality, and evaporation conditions.

3. To define the relationships between evaporator tube scaling, feedwater quality, and evaporation conditions.

THEORETICAL BACKGROUND

Volatile Organics

Treated sewage contains materials that can be classified as volatile under temperatures likely to be encountered in evaporation. These include numerous organic compounds. It would be desirable to predict how these will affect distillate quality. If sufficient data were available, Henry's law could be applied to make the prediction. Unfortunately, the limited work that has been done to characterize sewage treatment plant effluent plus the day-to-day variations in effluent quality make it virtually impossible to construct a predictive model for specific compounds in the effluent.[11,12] To complicate matters further, it appears that some of the organics in the evaporator feed are broken down by heat in the evaporation process to produce smaller, more volatile compounds.[7] This appears to be particularly true as evaporator temperature is increased. Therefore, it is impossible to predict either the amount or the specific compounds one might expect to carry over during evaporation.

Ammonia

Nitrogen in the form of ammonia is found in significant amounts in the effluent from almost all biological treatment processes. The exception is treatment involving extended oxidation conditions wherein the ammonia is oxidized to nitrate. Even in the extended aeration processes it is not unusual to find 0.5-1.0 mg/l NH_3-N.

The 1962 Public Health Service Drinking Water Standards[13] set no limit on ammonia concentration. Likewise the more stringent Water Quality Goals[14,15] established by the American Water Works Association make no mention of ammonia. In a 1968 report issued by the Federal Water Pollution Control Administration on raw water quality criteria for public supplies, a permissible criteria of 0.5 mg/l NH_3-N and a desirable criteria of less than 0.01 mg/l NH_3-N were recommended.[16] The rationale for these low limits was given as follows. "Ammonia is a significant pollutant in raw water for public water supplies because its reactions with chlorine result in compounds with markedly less disinfecting efficiency than free chlorine. In addition, it is frequently an indicator of recent sewage pollution."[16] In order to satisfy a chlorine demand of 1 mg/l NH_3-N and produce a free chlorine excess, i.e. "breakpoint" chlorination, about 10 mg/l of chlorine are required.[7] Hence, for economic reasons alone, it appears undesirable to have ammonia above 1.0 mg/l in the final product water. The fact that no standard for ammonia in potable water exists probably reflects that historically ammonia has not been a problem and not that ammonia was not considered undesirable.

The volatilization of ammonia in an ammonia-water system is influenced by pH, ammonia concentration, and solution temperature. Ammonia dissolves in water giving the following reaction:

$$NH_3 + H_2O \rightleftarrows NH_4^+ + OH^- \qquad (1)$$

$$K_b = \frac{[NH_4^+][OH^-]}{[NH_3]} \qquad (2)$$

If values of K_b at various temperatures are known, the relative concentrations of NH_4^+ and NH_3 can be calculated as a function of pOH. Now, since

$$K_w = [OH^-][H^+] \qquad (3)$$

and values of K_w at various temperatures are known, the relative concentrations of NH_4^+ and NH_3 as a function of pH can be determined.

$$[H^+] = \frac{[NH_4^+]\ K_w}{[NH_3]\ K_b} \qquad (4)$$

$$pH = \log \frac{[NH_3]}{[NH_4^+]} + pK_w - pK_b \qquad (5)$$

Values of pK_w[17,18] and pK_b[19,20] for ammonia are shown in Table 1. Using these values and equation 5, the distribution of NH_3 and NH_4^+ was calculated as a function of pH at various temperatures. Figure 1 shows that at 20°C and pH 7 more than 99% of the ammonia is present as NH_4^+ and hence not subject to evaporation. However, as the temperature increases, the equilibrium shifts and there is more ammonia available for evaporation. This would explain the fact that O'Connor[7] found that he had to lower the feed pH to under 5 in order to eliminate ammonia in the product water.

The relative volatility of ammonia as compared to water strongly influences the relative amounts of ammonia and water in the liquid and gas phases during evaporation. At the very low ammonia concentrations encountered in treatment plant effluents Henry's law should apply. Unfortunately, no data have been published for ammonia-water systems in the 0 to 20-mg/l NH_3-N concentration range. However, considerable data are available for ammonia-water systems in ammonia concentrations above 1% (10,000 ppm).[21-23] These data show the relative concentration of ammonia in the vapor phase to be much higher than in the liquid phase. This concentration factor increases as ammonia concentration in the liquid decreases and is around 10 at the 1% (ammonia in liquid) level. Therefore, at the much lower ammonia concentration levels we are considering, the ammonia concentration factor from liquid to gas should substantially exceed 10. Concentration of ammonia by evaporation and

TABLE 1

DISSOCIATION CONSTANTS FOR AQUEOUS AMMONIA AND WATER

Temp. °C	pK_w for water[17,18]	pK_b for Aqueous Ammonia[19,20]
0	14.9435	4.862
10	14.5346	4.804
20	14.1669	4.767
25	13.9965	4.751
30	13.8330	4.740
40	13.5348	4.730
50	13.2617	4.723
60	13.0171	---
100	12.30	4.87

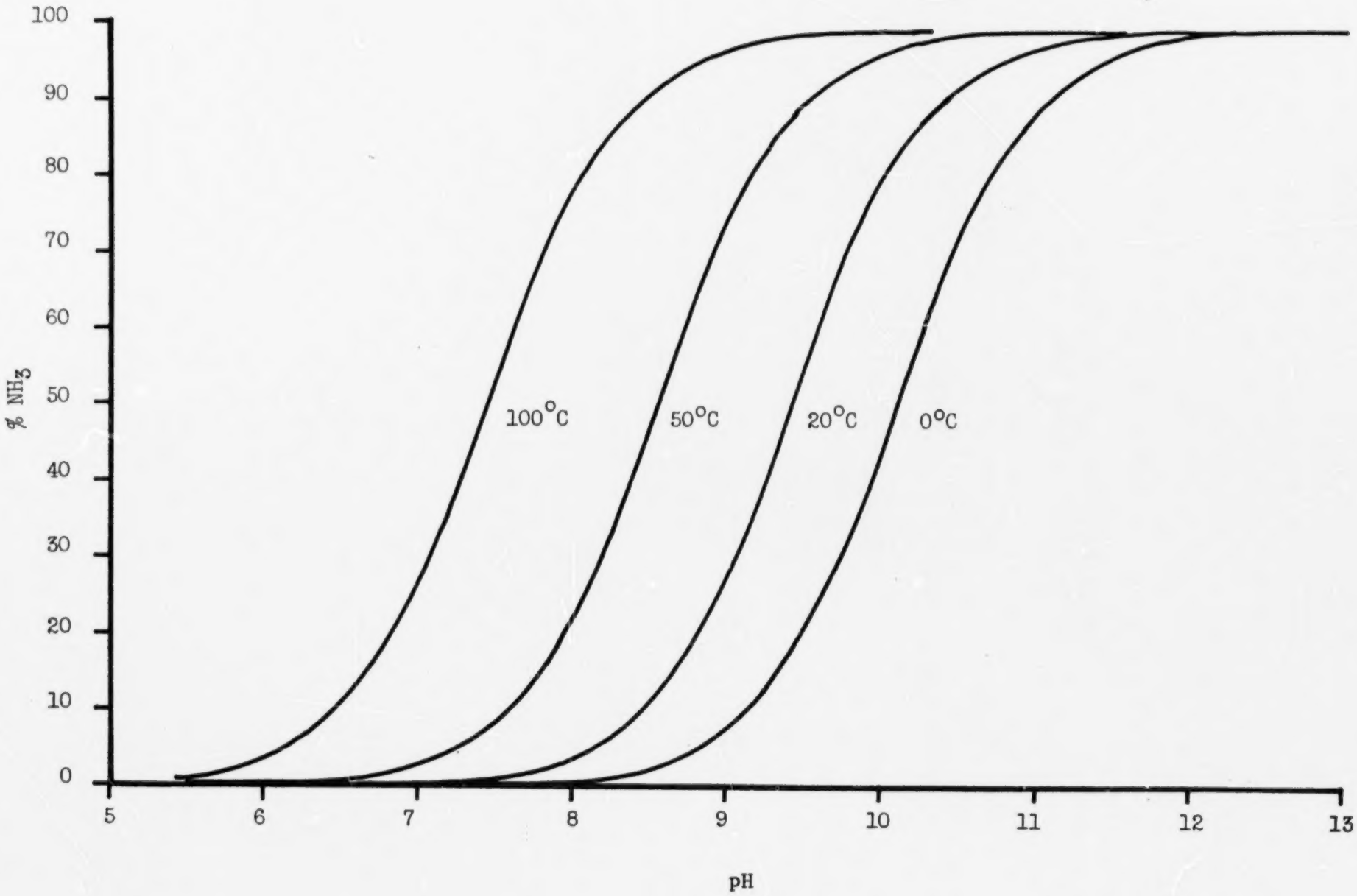

Figure 1. Distribution of NH_3 and NH_4^+ as a Function of pH and Temperature

condensation is well recognized and is an accepted laboratory procedure in the analysis of ammonia.[24]

The effect of temperature on ammonia volatilization from an ammonia-water system is that the ammonia concentration factor decreases slightly as temperature increases.[21] However, this effect does not appear to be significant over the temperature range that would be considered for evaporator operations.

From a theoretical standpoint, it appears that ammonia carryover to the product is dependent on the pH of the feed solution and the evaporation temperature. Since the ammonia-ammonium ion equilibrium shifts in favor of ammonia as temperature increases, one might well expect to find significant ammonia in the product water at pH values at least as low as 6 if evaporation is occurring under pressures exceeding atmospheric.

Scaling and Fouling

Loss of heat transfer efficiency may result from inorganic scaling, organic fouling, or a combination of the two.

Inorganic scaling occurs when the solubility of the scaling compound is exceeded. This may occur because of temperature increase, since the solubilities of many scale forming salts decrease with increasing temperature. In an evaporator, the continual evaporation of water concentrates the non-volatile ions in solution to the point that the entire solution may be saturated with respect to a particular salt. However, scaling may occur at localized points in the system even though the solubility of the particular scaling salt is not exceeded in the bulk of the water. This scaling can result from evaporation in the film of liquid immediately adjacent to the heating surface or merely from the higher temperature of the liquid in that film. The latter can occur only when solubility decreases with increasing temperature.

In the first case, where the entire solution becomes super-saturated, precipitation may occur throughout the bulk liquid. This precipitate may remain suspended in the solution and be removed with the concentrated effluent. However, it may settle on the tube surfaces and "bake" in place as a scale. The second case is more detrimental in that the precipitation tends to occur directly on the heating surface and there is little chance that the solid thus formed will be suspended in the solution.[25,26]

With regard to boiler scaling, the Betz Handbook[25] makes the following comments:

> "It is well to realize that the prevention of boiler scale cannot be predicted by any basic chemical principal. It is the physical characteristics of the precipitate formed in the boiler water that determines whether or not the precipitate will tend to tightly adhere to the boiler heating

surfaces - the chemical characteristics of the precipitate are relatively unimportant. Thus, from a chemical standpoint sodium silicate will precipitate the calcium and magnesium salts just as well as a phosphate or carbonate. However, it is the physical characteristics of the precipitate that are of importance, for a precipitate of calcium silicate will tightly adhere to the heating surfaces."

Treatment for prevention of inorganic scale may be external or internal. External treatment involves removing the potential scale forming materials from the water prior to its introduction into the evaporator, boiler, etc. Internal treatment involves the selective precipitation of the potential scale forming materials in the form of non-scaling sludges. This may include the addition of materials designed to keep the sludges fluid and in suspension so that they can be removed from the boiler or evaporator in the blowdown or concentrated effluent. The principal inorganic scale and/or sludge formers are calcium carbonate, calcium sulfate, magnesium hydroxide, and silica.[27]

Calcium Carbonate

Calcium carbonate is one of the least soluble salts that is likely to occur in wastewater evaporation. At temperatures below boiling this scale can be controlled by adjusting the carbonate equilibrium system in favor of H_2CO_3 and HCO_3^- with pH control. Excellent discussions of this equilibrium system have been presented by Langelier,[28,29] Dye,[30,31] and Weber and Stumm.[32] However, pH control of the carbonate system cannot be accomplished under boiling conditions where the CO_2 solubility is very low. Under such conditions HCO_3^- converts to $CO_3^=$ and H_2CO_3, and $CaCO_3$ can precipitate according to the following reactions.

$$2HCO_3^- \rightleftarrows H_2O + CO_2 + CO_3^= \qquad (6)$$

$$Ca^{++} + CO_3^= \rightleftarrows CaCO_3 \qquad (7)$$

Since CO_2 solubility is very low, the continual removal of CO_2 from the system forces essentially complete decomposition of the HCO_3^-.

One obvious way of controlling calcium carbonate scale is by removal of the calcium. This could be done by hot or cold lime-soda softening or by ion exchange.[25,27,33] However, in the case of wastewater distillation, these processes would add significant additional costs to the overall system. In addition, ion exchange is questionable because the effects of the organics on fouling of the exchange resin are not well known.

Calcium carbonate scale also can be controlled by pH adjustment and degassing prior to evaporation.[2] This involves lowering the pH to put the carbonate in the H_2CO_3 form and then removing it as CO_2 by degassing.

Finally, calcium carbonate scaling may be controlled internally by controlled precipitation of the calcium as a non-scaling sludge.[25,27,33] The most common procedure is to precipitate the calcium as hydroxyapatite, a mixed calcium phosphate and hydroxide. Several sodium phosphates may be used including trisodium phosphate, disodium phosphate, sodium metaphosphate, and monosodium phosphate.[25] In some cases it is necessary to add organic dispersants to prevent the sludge from growing into aggregates. Typical types of organic materials used for this purpose are tannins, lignins, glucose derivatives, and starches.[25,27,33]

Precipitation of calcium by phosphate and dispersal of the sludge by organics has promise for use in wastewater evaporation since both phosphates and organics naturally occur in this water. However, individual wastewater streams may or may not contain sufficient phosphate and the dispersal abilities of the organics in wastewater are not known.

Calcium Sulfate

Calcium sulfate presents a more serious problem than calcium carbonate because it forms a very hard and adherent scale.[27,33] Calcium sulfate solubility decreases with temperature as shown in Figure 2.[34,35] It should be noted that this is the solubility of pure calcium sulfate, anhydrite, in pure water. In wastewater, calcium sulfate will have a higher solubility due to the increase in ionic strength caused by the other ions in solution.

Control of calcium sulfate scale is accomplished in much the same way as control of calcium carbonate. External treatment to remove or reduce the calcium content of the water is applicable.[25,27,33] Internal treatment to remove calcium by controlled precipitation with phosphate can likewise be used.[25,27,33] The comments made previously regarding these control measures all apply here.

Magnesium Hydroxide

Magnesium hydroxide has a solubility of about 5 ppm at 212°F decreasing to slightly less than 1 ppm at 392°F. Control of this scale may be external by removal of the magnesium by softening or ion exchange.[25,27,33] Internal control can be accomplished by controlled precipitation with phosphate. However, magnesium phosphate tends to form a sticky deposit and requires mechanical removal.[25] Finally, since magnesium forms a hydroxide scale, it can be controlled by maintaining the pH and hence hydroxide concentrations at a low enough level so as not to exceed the solubility.

Silica

Silica tends to form complex scales which may include calcium, magnesium, aluminum, or may be composed almost entirely of silica.[25,27] These scales are usually very hard, glassy, and adherent. McCoy[33] gives a maximum permissible concentration of silica of 250 ppm at 388°F for

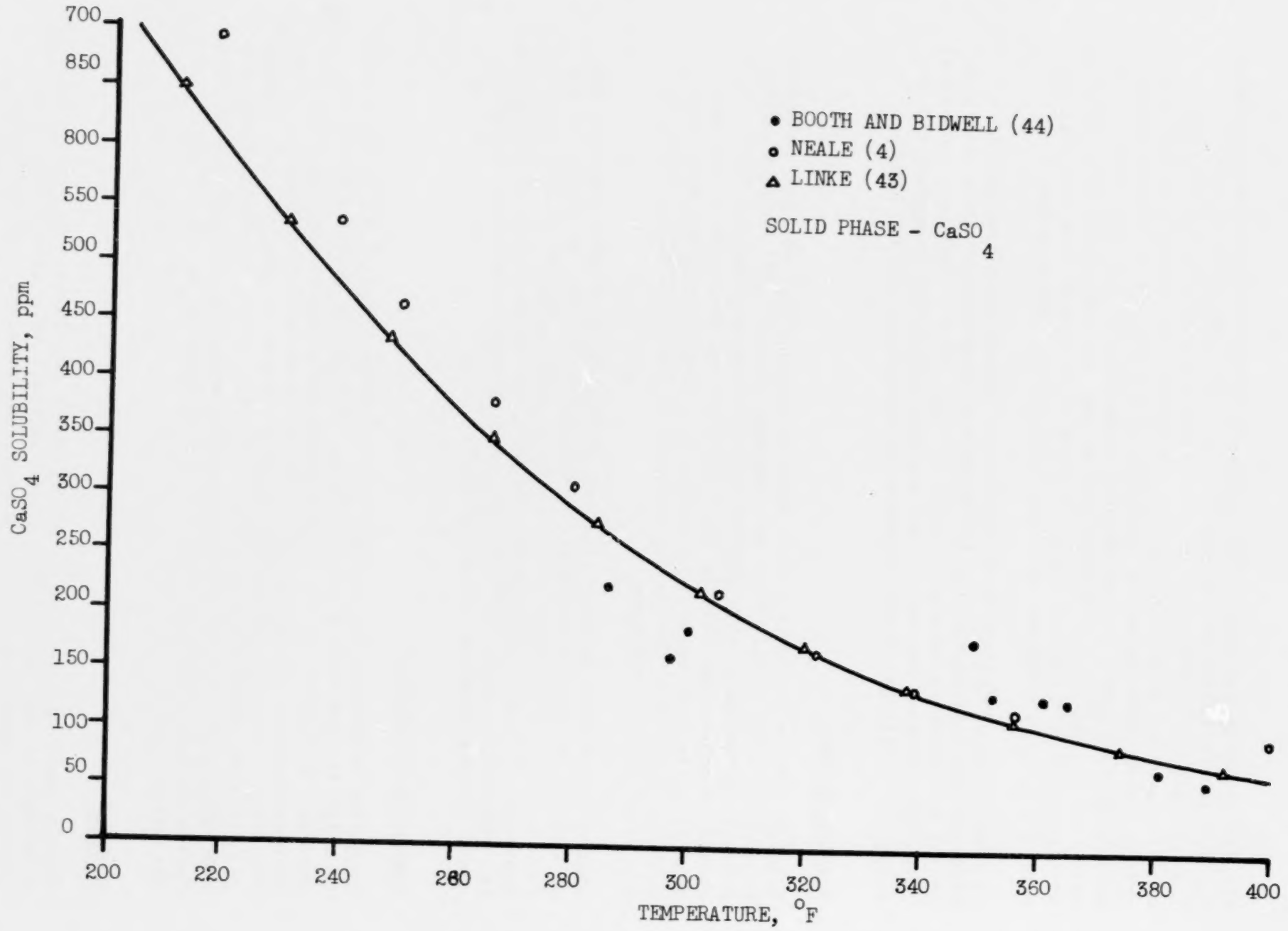

Figure 2. Calcium Sulfate Solubility

boilers. This figure drops to 175 ppm at 421°F. Presumably it increases for temperatures below 388°F.

In some cases too low silica concentrations can cause problems.[25] If magnesium concentration is high and phosphate is present, the magnesium will precipitate a sticky magnesium phosphate sludge. However, if sufficient silica is present, magnesium silicate, which is easier to handle, will be precipitated.

Control of silica scale may be external by hot lime-soda softening or by coagulation with ferric sulfate.[27,33] Ion exchange using a strongly basic anion exchange resin regenerated with sodium hydroxide may also be used.[27]

Organic Fouling

Organic materials may cause heat transfer problems by any of several different mechanisms. Suspended organics may simply deposit on heat transfer surfaces. Oils or greases, which have a tendency to coat metal surfaces, may cause problems. The high temperature in the evaporator may tend to polymerize some organics into high molecular weight insoluble compounds that could coat tube walls. In the latter stages of evaporation, as concentrations of both organic and inorganic materials increase, the possibility of salting out exists.

Another potential organic fouling problem exists in the early stage preheaters (temperature less than 130°F). This is the problem of bacterial growth. It is possible that this could be adequately controlled by periodic shock chlorination.

In contrast to all the potential problems with organic fouling, there are some possible positive benefits. Certain organics may have a dispersing effect which would tend to keep solids suspended. Also, volatile organic vapors carrying over from one effect to the steam chest of the next effect may promote dropwise condensation on the evaporator tubes. If the organic material coats the condensing surface and renders it non-wettable, condensation will occur dropwise and since dropwise condensation gives higher heat transfer, the overall efficiency of the process would be improved. [36]

Bacteria or Virus Contamination

Since in evaporation of sewage treatment plant effluent it is inevitable that bacteria and virus will be present in the feedwater, a question that must be faced is what are the possibilities of these organisms contaminating the product water. Three factors affect bacterial and viral contamination of the distillate. Considered together, these factors make contamination very unlikely. First, the organism would have to survive the maximum feed temperature in the evaporator. The forward feed multiple-effect evaporator design requires that <u>all</u> feedwater pass from the hot end of the system to the cold end. Hence, all incoming water is heated uniformly to the maximum operating temperature. In the case of seawater

this temperature is 250°F. However, as pointed out previously, it is likely that the upper temperature limit for sewage treatment plant effluent will be higher than this. For comparison purposes, milk is pasteurized at 161°F and microbiological laboratories sterilize their equipment in autoclaves at 250°F.[37] Sterilization times as low as 10 minutes are used for small samples. Because of the uniform heat conditions obtained in an evaporator, the somewhat shorter holdup times at the maximum temperature should be at least as efficient as 10 minutes in an autoclave.

The second factor dictating against bacterial or viral contamination of the product is the basic separation mechanism being used, i.e. evaporation and condensation. Since bacteria and virus are nonvolatile, they cannot contaminate the product by evaporating and condensing with the product. To get into the product water it would be necessary for any organisms to be carried over physically in the vapor stream, i.e. entrainment. The extent to which this would be possible would depend on the type and design of the vapor liquid separation system in each effect.

The third factor that would be important in eliminating any danger from bacterial or viral contamination would be chlorination. As with any municipal potable water supply, all water from the plant should be chlorinated.

Considering these facts, it seems reasonable to conclude that possible danger from bacteria or virus in a forward feed multiple-effect evaporator system using sewage treatment plant effluent as feed could be less than the same danger from a conventional coagulation plant using surface sources of raw water.

Ultimate Disposal

Practically all wastewater treatment processes involve separation of the incoming liquid into a "clean" and "dirty" stream. In all cases some final disposition must be made of the "dirty" or concentrated waste stream. This problem is very significant in evaporation of sewage treatment plant effluent. The final concentrated effluent stream from the evaporator probably will be on the order of 3-10% by volume of the incoming stream and contain essentially all the inorganic salts and organic material in the feed stream. Ultimate disposal of the stream will depend on the individual situation; however, several relatively simple means would be ocean outfall, underground disposal, or solar evaporation. When these methods cannot be used, more sophisticated and expensive techniques would have to be employed.

Because of the high temperature anticipated for wastewater evaporation, the concentrated effluent should be pathogenically safe for ocean disposal in a well-designed outfall. Since the effluent would have an appreciable organic content, provision should be made in the outfall design to release the waste far enough offshore so as to preclude the

possibility of contamination of beach or estuary areas. This method of disposal is becoming less popular and is almost certain to be more closely regulated in the future.

In areas where contamination of ground water would not be a problem, underground disposal of the concentrated effluent could be considered.

Climatic conditions and availability of land would determine whether or not disposal by solar evaporation would be possible.

Finally, it should be noted that ultimate waste disposal problems in an area would not be increased by evaporation of sewage treatment plant effluent. If evaporation were not used, the more dilute waste containing the same absolute amounts of impurities would still have to be disposed of. Therefore, it is likely that evaporation would reduce rather than add to ultimate disposal problems. If methods of disposal involve dilution, then the problem would be increased.

RESEARCH PLAN

The objectives of this project, particularly with regard to scaling evaluations, dictated that the work be done in equipment of sufficient size to simulate full scale operation, that operation be continuous rather than batch, and that various effluents typical of what could be expected from municipal waste treatment plants be used as evaporator feed. Within these constraints the following approaches were taken to various parts of the overall research plan.

Experimental Equipment

The long tube vertical (LTV) evaporator design was selected for use in this work. In discussing all types of evaporators, Standiford[38] points out that more evaporation is accomplished in LTV evaporators than any other type and makes the following comments regarding the LTV design.

> The widespread use of the LTV evaporator is due partly to the ability to build large single units, partly because of the high heat-transfer performance exhibited under most conditions, and partly because of the simplicity and hence cheapness of construction (simply a shell and tube heat exchanger surmounted by a vapor-liquid separator).
>
> Because of their high capacities, and lower costs than for any other type, rising-film LTV evaporators are used whenever possible . . . While they cannot usually handle crystallizing solutions, they are widely used for viscous and mildly scaling liquors. They are well suited to corrosive solutions because heat-transfer coefficients are generally high requiring a minimum of expensive heating surface, and tube replacement is simple . . . The principal disadvantage of the rising-film LTV evaporator is the poor heat-transfer performance at low temperature-differences or at low temperatures.

For these reasons and because the rising-film LTV design is readily adaptable to pilot scale size units, this type evaporator was selected for use in this research. Consideration was given to the use of a single-effect unit. It was concluded, however, that since any full-scale plant would be multiple-effect, pilot plant data from a multiple-effect unit would be far more valuable than data from a unit that would only simulate the first stage of a multiple-effect system.

Since corrosion was not to be studied in this research, it was deemed desirable to construct the evaporator of 316 stainless steel to minimize the possibility of product contamination by corrosion.

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As with the evaporator, the accessory equipment used for feed preparation and product polishing was designed to operate continuously.

In design of the evaporator provision was made to allow operation as a single stage unit or to operate all three effects in series. In either mode of operation any pressure range from about 28 in. Hg vacuum to 100 psig could be selected. Unfortunately, steam pressure available at the site limited the upper pressure to 50-55 psig.

The evaporator was constructed by Indian River Construction Co., Jacksonville, Florida in cooperation with Reynolds, Smith, and Hills, Architects and Engineers, Jacksonville, Florida.

Effluents Available

The experimental evaporator was set up adjacent to the 2-MGD Campus Sewage Treatment Plant operated by the University of Florida. This plant handles the flow from the main campus, the teaching hospital, and medical center complex, fraternities, sororities, dormitories, an elementary school, and several married student apartment complexes. Flow from all these sources results in a waste reasonably typical of largely domestic municipal sewage. The Campus Plant operates a primary treatment unit, a high rate trickling filter, a standard rate trickling filter, and a contact stabilization activated sludge unit. A 9,000-GPD extended aeration unit is also available for use but is not operated routinely. Effluents from any of these units are available for experimental purposes.

Experimental Plan

The experimental work was divided into two principal phases. The first phase was directed towards defining the relationships between feedwater quality, product quality, and evaporation conditions. Experimental runs were to be made operating the evaporator as a single effect unit and using progressively poorer quality effluents as feedwater. By this procedure the possibility could be minimized of having major difficulties with excessive organic fouling as was experienced during the work done at American Machine and Foundry, Inc.[3-6] Thus testing was to proceed using, in order, extended aeration plant effluent, contact stabilization plant effluent, high rate trickling filter effluent, and primary effluent until either operational difficulties or product water quality dictated that further testing with poorer quality effluents was not justified. Evaporator operating conditions were to be varied over the range from about 28 in. Hg vacuum to about 50 psig. Also, the pH of the feed was to be varied to see the effects on ammonia carry-over of the product. In addition to measuring the relationships between these variables and product water quality, the first testing phase included the evaluation of several post treatment processes. These included activated carbon treatment, aeration, and ammonia removal by ion exchange.

After completion of the first phase of the experimental work some preliminary judgements were to be made regarding the practical limits for both minimum feedwater quality and maximum evaporator operating temperature consistent with acceptable product water. Using conditions at or near these limits, extended continuous runs utilizing all three evaporator effects were to be made to evaluate scaling or fouling. It was anticipated that these runs would last one to two weeks each. Scaling was to be evaluated both directly and indirectly. Indirectly, scaling could be detected by decrease in overall heat transfer coefficient, U_o, for each stage. Direct measure of scaling was to be obtained by cleaning the evaporator mechanically and weighing the amount of scale removed. The scale was to be analyzed to determine its cause.

In addition to providing data on scaling, the extended runs would provide considerable additional data on the quality of product that could be expected under the conditions felt to be most likely for full scale operations.

Analytical Plan and Procedures

The analytical work involved three types of waters, sewage treatment plant effluent or evaporator feedwater, evaporator product water, and the concentrated effluent from the evaporator. Analyses were also conducted on evaporator scale.

Table 2 lists the analyses made on the liquid samples and the procedures used. The very large number of samples involved dictated that automated analytical techniques be used as much as possible.

Chemical Oxygen Demand, COD, was used as an indicator of the organic content of the samples. An attempt was made to perfect a total organic carbon procedure that would be accurate in the 0-3 mg/l range, the range anticipated for product samples. Unfortunately, with the equipment available, it was not possible to develop a procedure as precise as the dilute COD procedure and this effort was finally abandoned.

Ammonia nitrogen was one of the most important variables in this study. Its significance has been discussed previously.

Nitrate nitrogen was used to indicate the degree of nitrification in the various sewage treatment plant effluents. Another important use of nitrate results was to measure the degree of physical carry-over of material into the evaporator product water. Since nitrate is non-volatile, a material balance over the evaporator based on nitrate should be a very accurate measure of physical carry-over.

The importance of pH in ammonia distribution and in scaling has been discussed previously. Measurements were made primarily on feedwater samples. Samples were adjusted to $25^{o}C \pm 5^{o}C$ prior to measurement.

Odor was subjectively measured by the operating personnel at the time the products were produced. The difficulties involved in obtaining an odor

TABLE 2

ANALYTICAL TESTS ON LIQUID SAMPLES

Test	Procedure
COD	Alternate Procedure for Dilute Samples, Standard Methods for the Examination of Water and Wastewater[24]
Ammonia Nitrogen	Industrial Method Ind.-19-69w of Technicon Corp. Used Technicon AutoAnalyzer
Nitrate Nitrogen	Method of Kahn and Brezenski[39] modified for Technicon AutoAnalyzer
pH	Corning Model 12 pH meter
Odor	Subjective opinion of operator
Total Dissolved Solids	Residue on Evaporation at 103°C, Standard Methods for the Examination of Water and Wastewater[24]

free room and an experienced panel dictated against more rigorous odor testing.

Total dissolved solids were measured on a few of the feedwaters and concentrated effluent samples, primarily to better characterize the evaporator feedwater.

Some additional test results taken from the records of the Campus Sewage Treatment Plant were used to characterize influent and effluent materials from the various plant processes.

The scale removed from the evaporator was first dried at 103°C and weighed to obtain a total weight figure. The samples were then fired at 600°C to remove the organic material. The fired residues were dissolved in dilute hydrochloric acid and the solution of the soluble portion analyzed as shown in Table 3.

TABLE 3

ANALYTICAL TESTS ON SCALE

Test For	Procedure Used
Ca	EDTA Titrimetric Method, Standard Methods for the Examination of Water and Wastewater[24]
Mg	Atomic Absorption Method Using Beckman DB-G Spectrophotometer
Na	Same as above
Fe	Phenanthroline Method, Standard Methods for the Examination of Water and Wastewater[24]
SO_4	Turbidimetric Method, Standard Methods for the Examination of Water and Wastewater[24]
PO_4	Method of Murphy and Riley[40]

Note: Fired scale samples were dissolved in 0.5N HCl

DESCRIPTION OF EQUIPMENT

A general flowsheet for the process is shown in Figure 3. Feedwater for the system could come from either an extended aeration plant, a trickling filter plant, or a contact stabilization plant. Before any of these effluents were fed into the evaporator, the dissolved gases and carbonates were removed. This was accomplished by adjustment of the pH and degassing under a vacuum. After evaporation, provision was made to treat the product water either in an activated carbon column or an ammonia selective ion exchange column or both.

Degasser

A diagram of the degassing equipment is shown in Figure 4. Effluent from the various sources was put into a 175-gal. agitated tank for mixing and pH adjustment. The agitator was a flat blade 30 in. in diameter by 1 3/8 in. high and was driven at 100 rpm. From the mix tank, the effluent passed through a packed column under vacuum to remove dissolved gases. The packed column was 4-in.-O.D. glass tubing. Vapor and dissolved gases were removed from the column through a condenser and liquid trap by a sliding vane type vacuum pump. Degassed liquid was pumped from the bottom of the column to one of two 55-gal. open top feed drums by a variable speed screw type pump. The two feed drums were fitted with polyethylene drum liners. The mixing tank, piping, and pump in the degassing system were of steel construction.

Temperature in the column was measured by a thermometer inserted into the column through a small port. Provision was made to introduce steam into the bottom of the column.

Evaporator

A diagram of the evaporator is shown in Figure 5 and a picture of the evaporator is shown in Figure 6. The evaporator was designed to operate either as a triple-effect unit or as a single-stage unit (3rd effect only). Possible operating conditions in the evaporator ranged from about 28 in. of Hg vacuum to 100 psig. Actual operating conditions under pressure conditions were limited to about 55 psig by the steam pressure available at the site.

Heat for the evaporator could be supplied to either the first or third effect. The heating medium could be either steam or hot water. Steam was taken from the campus steam line and was reduced at the inlet to the evaporator by a steam regulator. When operating under high vacuum, it was necessary to reduce the temperature of the heating medium to avoid unrealistically high temperature drops. This was done by using a closed hot water system. The water was circulated continuously through the evaporator using a natural gas fired hot water heater as the heat source.

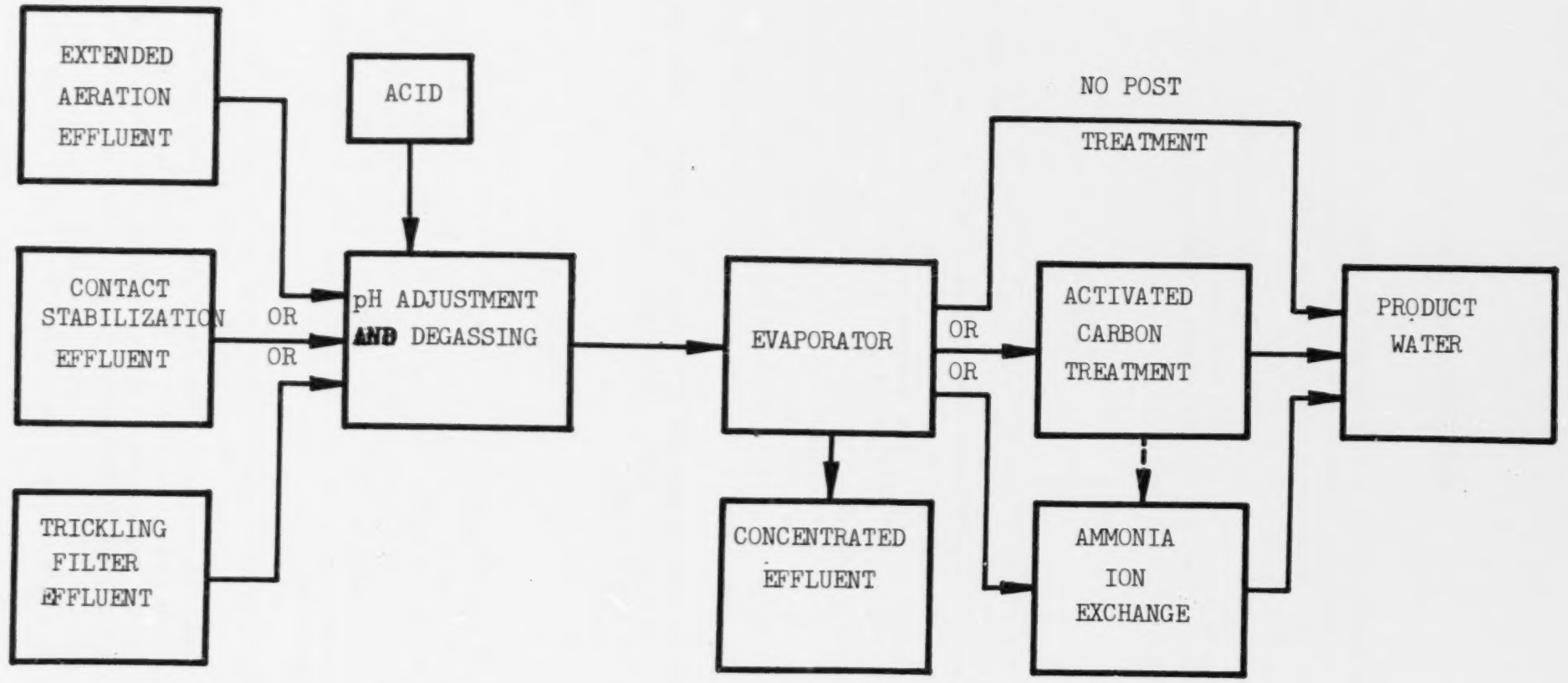

Figure 3. Overall Flowsheet

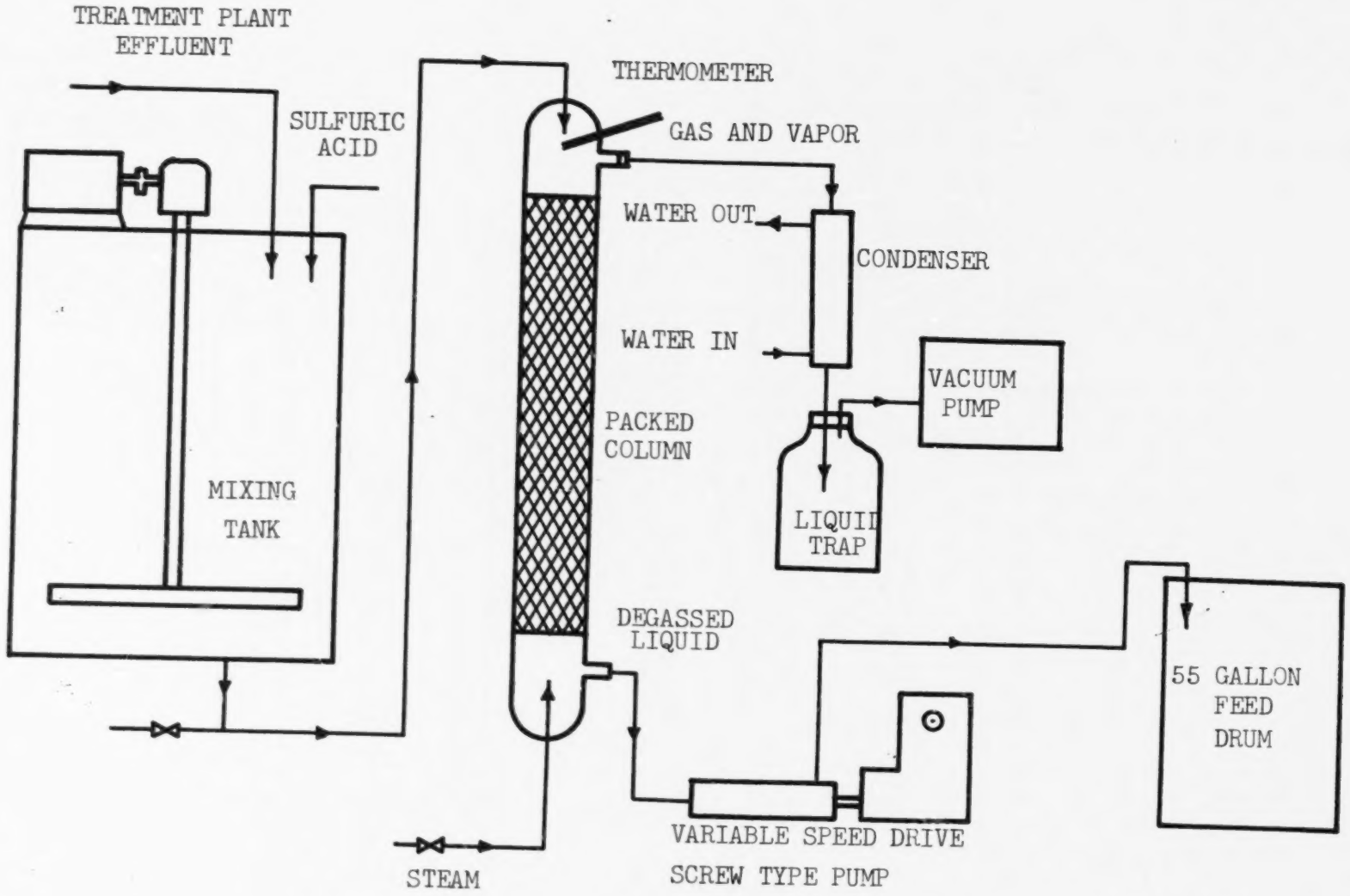

Figure 4. Degassing System Flowsheet

VLS - VAPOR LIQUID SEPARATOR
LLC - LIQUID LEVEL CONTROL
HWO - HOT WATER OUT
HWI - HOT WATER IN
CWO - COOLING WATER OUT
CWI - COOLING WATER IN
VAP.- VAPOR
LIQ.- LIQUID

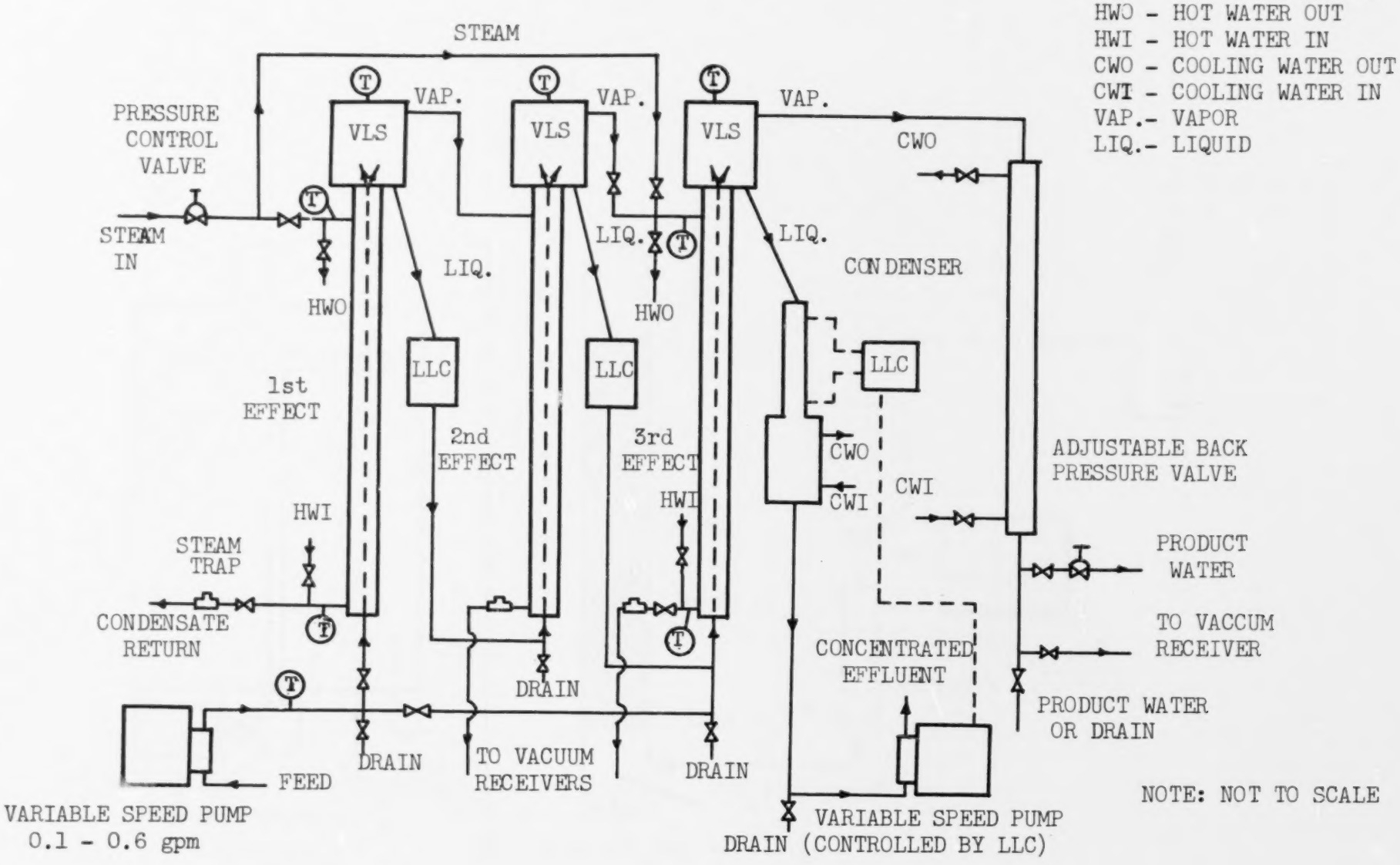

Figure 5. Evaporator Flowsheet

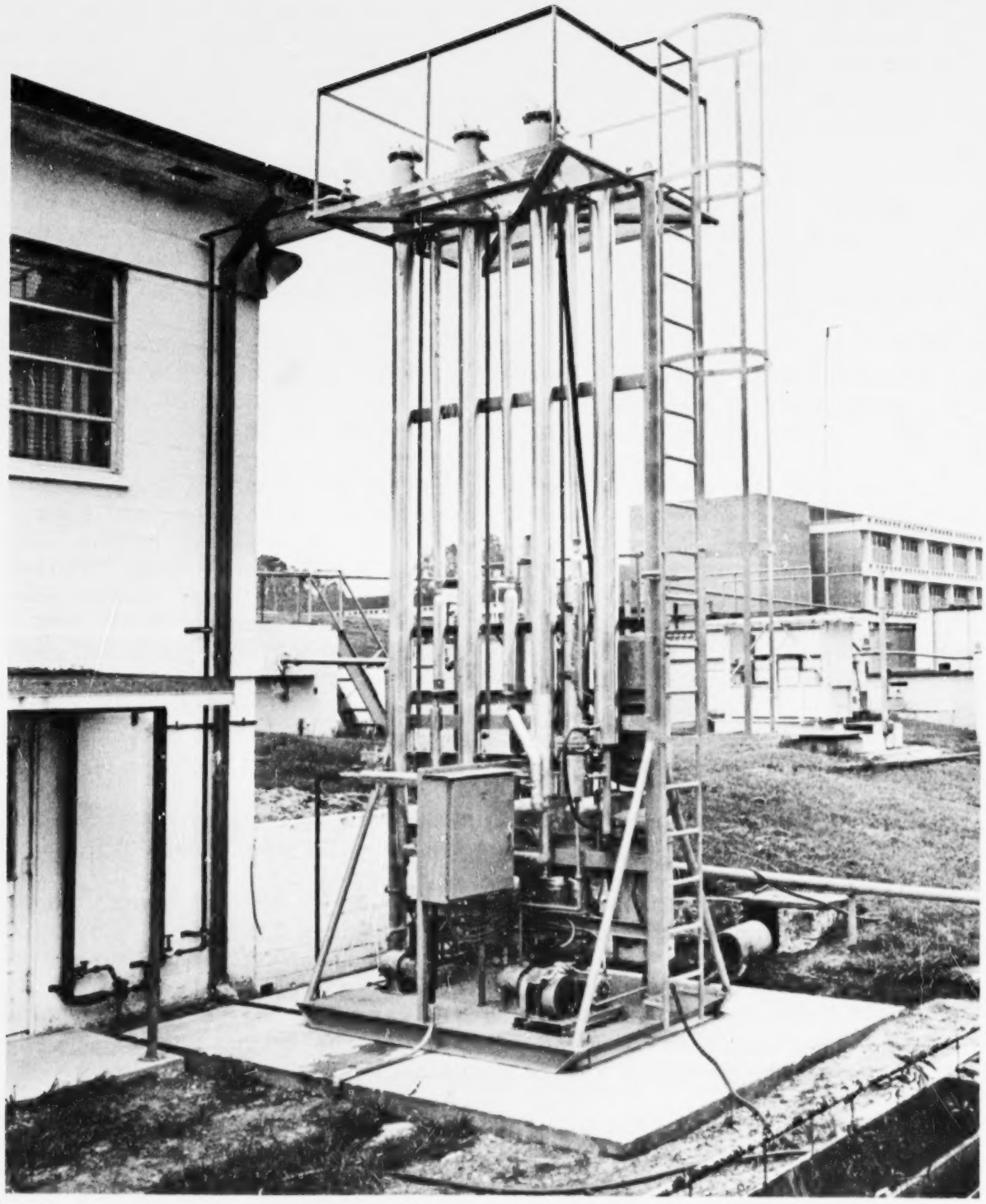

Figure 6. Evaporator

Feed was pumped into the evaporator by a diaphragm metering pump. A spring loaded, back pressure valve on the downstream side of the pump prevented liquid from being drawn through the pump when the system was operating under vacuum conditions. Each of three effects was constructed of 1-in. IPS Schedule 5 pipe, jacketed with 4-in. IPS Schedule 5 pipe. The effective heated length of each effect was 14 ft.

Vapor-liquid separators atop each effect separated the steam-liquid mix into two streams. This equipment is shown in Figure 7. The demisters were 316 stainless steel mesh (Otto H. York Co., Style 326). Small vent lines (1/4-in. O.D. tubing) with needle type control valves connected the steam jacket of each effect with its vapor-liquid separator.

Liquid seals were maintained between stages by float controlled liquid drainers having 7/32-in. diameter orifices. These drainers were chrome plated cast iron with stainless steel internals. Originally, drainers with 1/16-in. diameter orifices were tried, but were replaced because the capacity was marginal and chances of plugging were much greater than with larger units.

To reduce heat loss to the atmosphere, the three evaporator effects and connecting piping were insulated with 1-in. thick asbestos type insulation. The three vapor-liquid separators were completely enclosed with custom built foamglass type insulating covers with aluminum exteriors.

Removal of the concentrated liquid effluent from the last effect was accomplished by a diaphragm metering pump identical to the feed pump. This pump was automatically controlled (on-off) by liquid level probes located in an enlarged section of the liquid line from the third effect vapor-liquid separator. Cooling coils, located between the liquid level probes and the pump, cooled the concentrated effluent to prevent flashing of the hot liquid when released to atmospheric pressure. All concentrated effluent piping was steel. The back pressure valve following the pump was a copper alloy.

Condensate was removed from the jackets of the three effects through controlled disc type steam traps.

Vapor from the third effect was condensed in a water jacketed condenser. The condenser was a 2-in. IPS pipe with an effective condensing length of 12 ft.

Product from the three stages was removed by several means, depending on the evaporator operating conditions (See Figures 5 and 8). Under pressure conditions, product no. 3 was removed from the final condenser through an adjustable back pressure valve. This valve was of chrome plated copper alloy construction. Product Nos. 1 and 2 were removed through the vacuum receiver system as shown in Figure 8 except that the system was opened at valves D allowing the products to drain continuously to glass carboys.

Under vacuum operation all three products were removed through the vacuum receiver system. Under normal conditions, valves B and C were closed and valves A were open. Valves D were check valves that allowed flow from the upper to the lower tanks only. Hence, product water would drain from the evaporator into the upper tanks and down into the lower tanks through valves D. Vacuum was maintained on the system by a vacuum pump. In order to remove

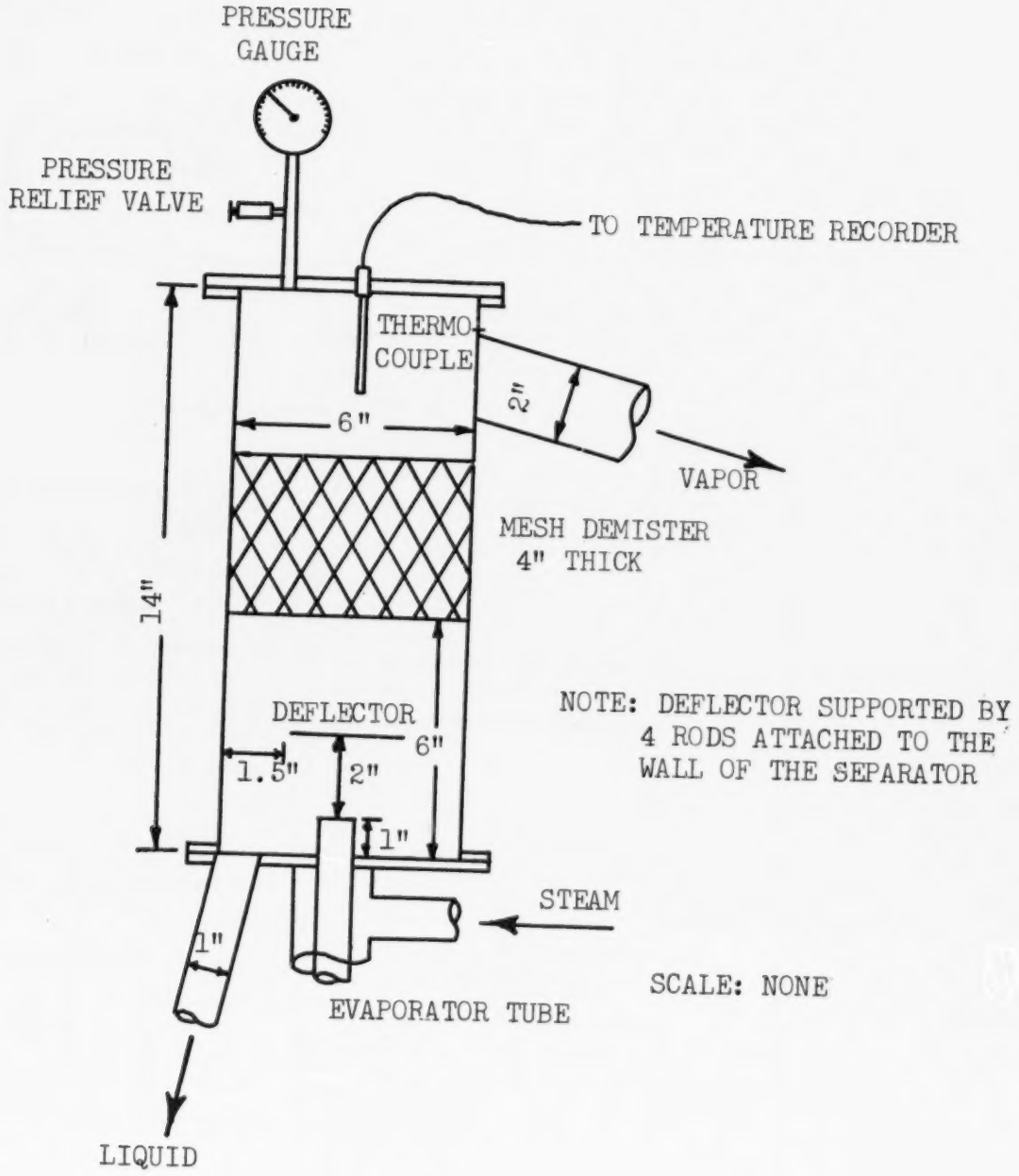

FIGURE 7. Vapor-Liquid Separator

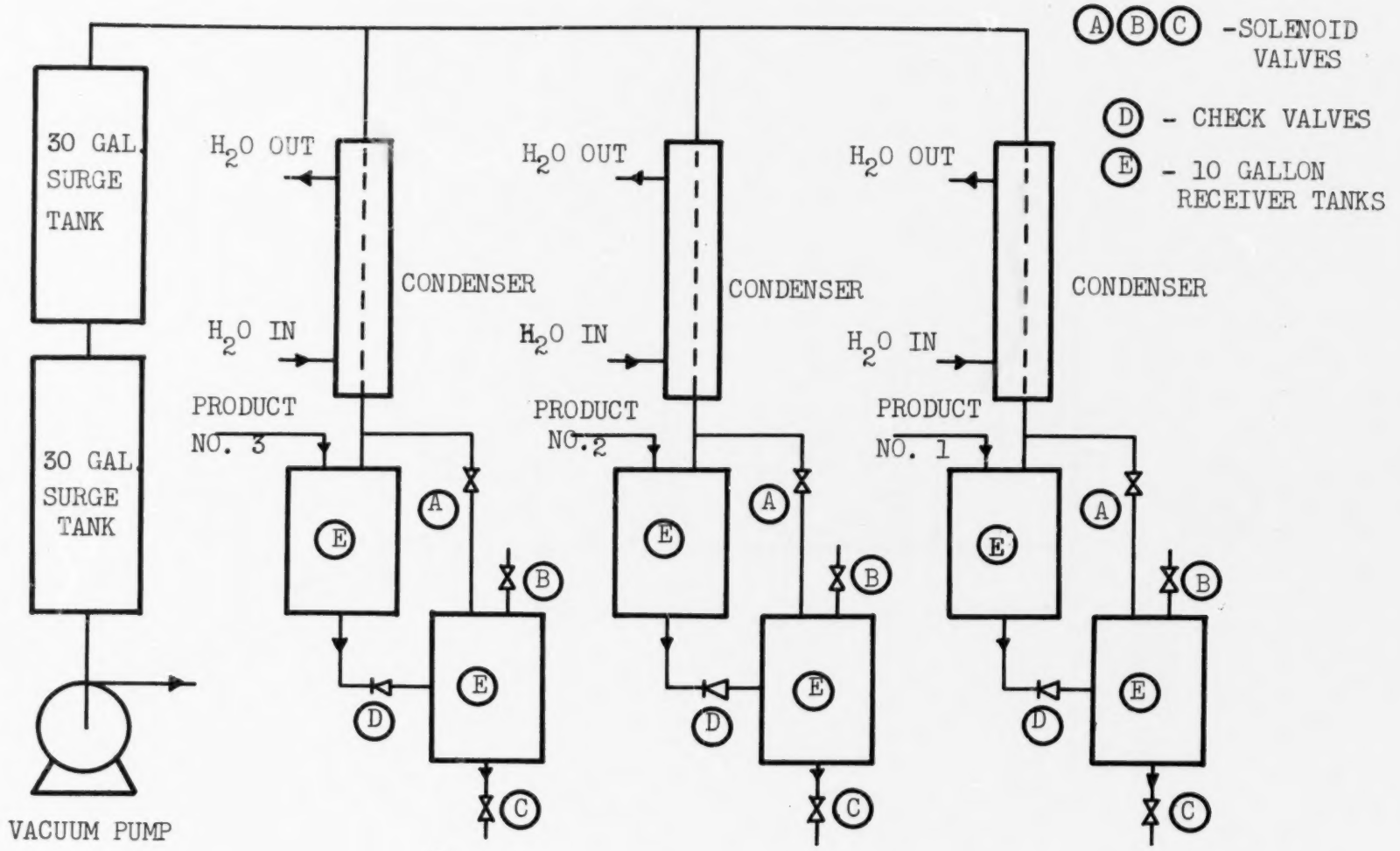

Figure 8. Product Receiver System Flowsheet

the product water from any one system, solenoid valves B and C were opened and valves A were closed. The product water either drained through valves C by gravity or was drawn into evacuated carboys. When the lower tank was empty, the cycle was reversed closing valves B and C and opening valves A. Two 30-gal. tanks in the vacuum system provided surge capacity when an air filled tank was cycled back into the system. The instrumentation system provided for automatic cycling of the receiver system to dump product water from each effect on a timed sequence.

All six product receiver tanks were 10-gal. capacity.

Temperature was sensed by copper-constantan thermocouples at various points in the system and automatically recorded by a 16 point recorder. The temperature range of this recorder was 0-350°F. Pressure gauges, installed in the incoming steam line, in each vapor-liquid separator, and in the final condenser, gave continuous visual indication of the pressure at various points in the system.

Carbon Column

The carbon column was a 1.94-in.-I.D. pyrex tube 4 ft. long, containing 42 in. of activated carbon. The column was fed by gravity and flow rate was controlled by a polyethylene needle valve. The activated carbon used was a Nuchar WV-W, 12 by 40 mesh, manufactured by West Virginia Pulp and Paper Co.

Ammonia Column

The ammonia ion exchange column was a 1.94-in. pyrex tube 13 in. long, containing 10.25 in. of ion exchange material. The column was fed by gravity and the flow rate was controlled by a polyethylene needle valve. The ion exchange material used was natural clinoptilolite, 20 by 50 mesh, from the Hector, California, area.

EXPERIMENTAL RESULTS

Initial Testing

As stated in the research plan, the initial testing was designed primarily to define the relationships among evaporator feedwater quality, evaporator operating conditions, and product water quality. Also, to be included was evaluation of several post treatment processes and determination of how they might be included in an overall reuse cycle.

With regards to effluents to be used, it was planned to begin the testing with the highest quality effluent to reduce the likelihood of severely scaling or fouling the evaporator tubes. Testing would then proceed using progressively poorer quality effluents.

The tests on each different effluent were made at three temperature and pressure levels. These conditions are shown in Table 4. In addition to the temperature variations, the pH values of the feed were varied at each level to examine the effect on ammonia distribution between the concentrated effluent and product.

On selected evaporator products that would be unsuitable for direct reuse, tests were made using activated carbon, ammonia selective ion exchange material, and aeration to determine the effectiveness of these post treatment processes.

Before any sewage plant effluents were used as evaporator feed, the evaporator was operated using tap water and distilled water. This was done to clean out the inside of the equipment and to develop familiarization with the operating characteristics of the equipment. Steam was used to clean the product receiver tanks.

When the evaporator was operated as a single-effect unit (3rd stage only), equilibrium conditions for the entire system could be reached in less than 15 minutes. The only exception occurred when the feedwater was changed during a run. When this occurred it could take as long as one hour to flush completely the system and obtain consistent quality product water.

In these initial tests the operating procedure was as follows: The effluent to be used was either pumped or drained by gravity from the sewage treatment plant to the mixing tank. A sample was taken for analysis and the pH of the remaining water was adjusted to the bicarbonate endpoint (pH 4.8-5.0) with 6N sulfuric acid. The water was heated to 120-140^{o}F and then degassed under vacuum. The degassed water was pumped to 55-gal. drums prior to being fed into the evaporator. If required, a pH adjustment of the evaporator feedwater was made at this point using either sulfuric acid or sodium hydroxide.

In the degassing operation there was a 3-5^{o}F drop in temperature caused by flashing as the water entered the top of the column. This corresponded to the measured 0.3-0.5% (of incoming water) evaporation rate from the degasser.

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TABLE 4

EVAPORATOR OPERATING CONDITIONS FOR
INITIAL SINGLE STAGE TESTS

Condition	Evaporating Liquid Temperature and Pressure	Heating Medium Temperature and Pressure
Vacuum	114-120°F 26.5-27 in. Hg Vac.	Hot Water 140-155°F
Atmospheric	212°F 0 psig	Steam 228-238°F 5-9 psig
Pressure	286-288°F 39-41 psig	Steam 300-305°F 52-58 psig

The liquid evaporated during degassing usually had a slight odor. However, the odor was normally no stronger than the odor of products produced under atmospheric pressure conditions.

The average degassing rate was 2800 ml/min but ranged from 2100 to 3900 ml/min. This corresponds to an average loading rate in the degassing column of 9.7 gpm/ft^2 and a range of 7.2-13.4 gpm/ft^2.

After evaporator start-up, from 30 to 90 minutes were allowed to attain stable operating conditions and completely flush the system. Product sampling was then begun, usually at 60-minute intervals. Since for each operating condition one drum of feed was required, only one feed sample was taken. One concentrated effluent sample was taken under each set of conditions. Total operating time, once product sampling began, was usually 2 hours, but some runs as long as 5 hours were made.

Extended Aeration Plant Effluent

Since extended aeration plant effluent was the highest quality secondary effluent available, testing was begun using this water. The effluent was from a 9,000-gpd-capacity plant built by Chicago Pump. This plant was fed raw sewage that had previously been degritted and ground. Flow to the plant was maintained at a constant rate by a constant head and weir arrangement. Excess raw sewage from the Campus Sewage Treatment Plant was pumped to a constant head headbox and the desired amount removed over a V-notch weir. The excess was returned to the Campus Plant. Flow to the plant during the period that the initial batch tests were being made was 6.2 gpm (8,920 gpd). The plant had been in operation for about four months prior to the initial tests and the mixed liquor suspended solids in the aeration basin had stabilized at about 4,200 mg/l.

Thirteen runs were made using extended aeration plant effluent, five each under atmospheric and pressure conditions and three under vacuum conditions. The results of these runs are shown in Tables 5, 6 and 7.

All products had odors strong enough to eliminate the possibility of direct municipal reuse. These odors ranged from very slight musty for products produced under vacuum to moderately strong fecal for products produced under pressure conditions.

Contact Stabilization Plant Effluent

Contact stabilization plant effluent was taken from the Campus Sewage Treatment Plant. This unit treats about 0.7 MGD of raw sewage. Performance data for this plant are shown in Appendix I.

Effluent from this unit was taken from the final clarifier prior to chlorination. Fourteen runs were made, five each under vacuum and pressure conditions and four under atmospheric conditions. The results of these runs are shown in Tables 8, 9, and 10.

TABLE 5

INITIAL TESTS, EXTENDED AERATION EFFLUENT,
VACUUM CONDITIONS

	Run No. 1	2	3
Operating Conditions			
Feed Rate, ml/min	647	647	632
Evaporation Rate, ml/min	93	95	117
% Evaporated	14.4	14.7	18.5
Hot Water Temp., ^{o}F	144	144	155
First Stage Temp., ^{o}F	116	116	122
Feed Temp., ^{o}F	91-96	98-106	95-103
Feed, Before Degassing			
Alkalinity, mg/l $CaCO_3$	25	25	32
pH	6.5	6.5	6.5
COD, mg/l	49	49	65
NH_3-N, mg/l	0.8	0.8	0.1
NO_3-N, mg/l	19	19	---
Feed, After Degassing			
pH	5.3	5.8	6.6
COD, mg/l	46	47	71
NH_3-N, mg/l	0.7	0.9	0.1
NO_3-N, mg/l	18	---	---
Product			
pH	6.6	5.9	7.2
COD, mg/l	1.3	2.6	4.5-6.3
NH_3-N, mg/l	0.04	0.11	<0.1-0.3
NO_3-N, mg/l	22	---	---
Concentrated Effluent			
pH	6.1	6.2	7.2
COD, mg/l	55	54	71
NH_3-N, mg/l	1.0	1.0	---
NO_3-N, mg/l	22	---	---

TABLE 6

INITIAL TESTS, EXTENDED AERATION EFFLUENT,
ATMOSPHERIC CONDITIONS

	Run No. 1	2	3	4	5
Operating Conditions					
Feed Rate, ml/min	640	630	630	655	640
Evaporation Rate, ml/min	156	146	177	186	293
% Evaporated	24.4	23.2	28.1	28.0	45.8
Steam Temp., ^{o}F	226	226	230	229	238
First Stage Temp., ^{o}F	212	212	212	212	212
Feed Temp., ^{o}F	96-104	106-114	88-104	90-104	80-82
Feed, Before Degassing					
Alkalinity, mg/l $CaCO_3$	27	27	34	40	50
pH	6.5	6.5	6.7	6.3	6.7
COD, mg/l	23	23	27	57	85
NH_3-N, mg/l	1.4	1.4	0.42	0.09	0.42
NO_3-N, mg/l	7.0	7.0	---	9.0	8.3
Feed, After Degassing					
pH	5.4	5.9	6.1	6.3	6.5
COD, mg/l	36	32	28	65	74
NH_3-N, mg/l	0.35	0.37	0.31	0.08	0.43
NO_3-N, mg/l	7.5	7.3	---	9.0	8.3
Product					
pH	5.5	5.7	5.9	6.0	6.1
COD, mg/l	0.0-0.6	0.5-0.7	2.6-3.2	3.8-5.0	1.8-5.8
NH_3-N, mg/l	0.05	0.10	0.16	0.03	0.36
NO_3-N, mg/l	0.00	0.00	---	---	---
Concentrated Effluent					
pH	6.3	6.4	6.5	6.3	6.7
COD, mg/l	41	40	37	87	137
NH_3-N, mg/l	0.44	0.47	1.8	0.10	0.38
NO_3-N, mg/l	8.1	8.1	---	13	14

TABLE 7

INITIAL TESTS, EXTENDED AERATION EFFLUENT,
PRESSURE CONDITIONS

	Run No. 1	2	3	4	5
Operating Conditions					
Feed Rate, ml/min	630	635	625	625	630
Evaporation Rate, ml/min	165	166	146	182	219
% Evaporated	26.2	26.2	23.4	29.1	34.8
Steam Temp., ^{o}F	299	299	300	302	304
First Stage Temp., ^{o}F	286	286	288	288	288
Feed Temp., ^{o}F	96-104	104-114	100-115	98-118	109-120
Feed, Before Degassing					
Alkalinity, mg/l $CaCO_3$	---	---	42	57	49
pH	6.5	6.5	6.5	6.5	6.7
COD, mg/l	52	52	97	110	85
NH_3-N, mg/l	1.0	1.0	0.40	1.3	0.42
NO_3-N, mg/l	7.7	7.7	9.3	7.4	8.3
Feed, After Degassing					
pH	4.8	5.6	5.9	6.2	6.5
COD, mg/l	47	48	90	94	74
NH_3-N, mg/l	1.0	0.8	0.33	1.4	0.43
NO_3-N, mg/l	7.9	7.7	8.6	7.0	8.3
Product					
pH	5.5	5.5	6.2	6.8	6.3
COD, mg/l	3.1-3.2	2.8	3.0-5.1	5.3-128	3.3-5.0
NH_3-N, mg/l	0.15	0.22	0.30	1.2	0.59
NO_3-N, mg/l	0.01	0.02	---	---	---
Concentrated Effluent					
pH	5.9	6.1	6.5	6.0	6.4
COD, mg/l	56	52	106	104	106
NH_3-N, mg/l	1.4	1.4	0.43	1.5	0.36
NO_3-N, mg/l	8.4	8.5	12	10	12

TABLE 8

INITIAL TESTS, CONTACT STABILIZATION EFFLUENT,
VACUUM CONDITIONS

	Run No. 1	2	3	4	5
Operating Conditions					
Feed Rate, ml/min	630	635	625	645	630
Evaporation Rate, ml/min	78	88	90	128	79
% Evaporated	12.4	13.9	14.4	19.8	12.5
Hot Water, Temp., ^{o}F	140	141	141	146	140
First Stage, Temp., ^{o}F	114	113	114	113	114
Feed Temp., ^{o}F	89-93	87-93	91-96	92-108	93-106
Feed, Before Degassing					
Alkalinity, mg/l $CaCO_3$	70	82	82	93	70
pH	6.9	7.1	7.1	7.2	6.9
COD, mg/l	23	23	23	30	23
NH_3-N, mg/l	5.1	6.5	6.5	9.0	5.1
NO_3-N, mg/l	---	0.9	0.9	0.28	---
Feed, After Degassing					
pH	5.6	5.6	6.3	6.7	7.7
COD, mg/l	29	20	19	35	24
NH_3-N, mg/l	6.4	6.8	6.6	8.5	5.8
NO_3-N, mg/l	0.5	0.4	0.5	0.31	---
Product					
pH	6.9	6.7	7.4	8.4	9.1
COD, mg/l	3.6	2.5	3.2	0.9-2.9	4.0
NH_3-N, mg/l	1.6[a]	0.14	0.9	2.9	6.9
NO_3-N, mg/l	---	---	---	0.00	---
Concentrated Effluent					
pH	7.4	6.8	7.3	7.4	8.2
COD, mg/l	29	22	22	42	23
NH_3-N, mg/l	6.4	7.7	7.3	7.6	5.9
NO_3-N, mg/l	0.6	0.4	0.5	0.36	---

[a] This run immediately followed Run No. 5 and the high NH_3-N result may have been due to contamination from the previous run.

TABLE 9

INITIAL TESTS, CONTACT STABILIZATION EFFLUENT,
ATMOSPHERIC CONDITIONS

	Run No. 1	2	3	4
Operating Conditions				
Feed Rate, ml/min	635	640	630	640
Evaporation Rate, ml/min	169	144	172	163
% Evaporated	26.6	22.5	27.3	25.4
Steam Temp., °F	230	227	230	229
First Stage Temp., °F	212	212	212	212
Feed Temp., °F	95-104	94-106	97-108	84-108
Feed, Before Degassing				
Alkalinity, mg/l $CaCO_3$	---	101	---	93
pH	6.9	7.1	6.9	7.1
COD, mg/l	24	68	24	27
NH_3-N, mg/l	3.3	6.6	3.3	7.0
NO_3-N, mg/l	3.1	0.17	3.1	0.49
Feed, After Degassing				
pH	5.1	5.7	5.9	6.5
COD, mg/l	29	62	25	32
NH_3-N, mg/l	3.5	---	4.3	9.2
NO_3-N, mg/l	3.1	0.23	3.1	0.48
Product				
pH	5.7	7.1	6.4	6.6
COD, mg/l	4.7-5.0	2.9-3.5	2.1-3.0	1.1-2.7
NH_3-N, mg/l	0.22	0.9	0.3	4.2
NO_3-N, mg/l	---	---	---	0.01
Concentrated Effluent				
pH	6.4	6.4	6.4	7.5
COD, mg/l	30	64	27	32
NH_3-N, mg/l	4.0	---	3.7	10.0
NO_3-N, mg/l	4.1	0.41	4.3	0.64

TABLE 10

INITIAL TESTS, CONTACT STABILIZATION EFFLUENT,
PRESSURE CONDITIONS

	Run No. 1	2	3	4	5
Operating Conditions					
Feed Rate, ml/min	635	630	630	615	630
Evaporation Rate, ml/min	152	160	173	212	170
% Evaporated	24.0	25.4	27.4	34.4	27.0
Steam Temp., °F	299	299	300	301	299
First Stage Temp., °F	286	286	287	284	286
Feed Temp., °F	94-103	86-96	102-111	91-104	97-108
Feed, Before Degassing					
Alkalinity, mg/l $CaCO_3$	91	69	91	88	69
pH	7.0	7.1	7.0	7.0	7.1
COD, mg/l	23	24	23	35	24
NH_3-N, mg/l	6.7	4.6	6.7	7.5	4.6
NO_3-N, mg/l	0.1	---	0.1	0.38	---
Feed, After Degassing					
pH	5.2	5.7	6.5	6.6	7.2
COD, mg/l	22	23	22	25	24
NH_3-N, mg/l	7.5	5.7	7.2	8.1	6.4
NO_3-N, mg/l	0.2	0.1	0.2	0.31	1.1
Product					
pH	6.3	6.0	6.9	6.6	6.6
COD, mg/l	1.7-2.5	4.2-4.7	4.0-4.9	4.8-5.2	1.5-3.4
NH_3-N, mg/l	2.2	4.5	3.3	3.5	9.4
NO_3-N, mg/l	---	---	---	0.02	---
Concentrated Effluent					
pH	6.8	8.3	7.3	6.8	6.8
COD, mg/l	24	28	25	39	27
NH_3-N, mg/l	8.5	5.3	7.8	8.9	4.0
NO_3-N, mg/l	0.3	0.9	0.4	0.55	1.3

All products had odors sufficiently strong to eliminate the possibility of direct municipal reuse. These odors ranged from musty for products produced under vacuum conditions to a rather strong, disagreeable, fecal type odor for products produced under pressure conditions.

Trickling Filter Plant Effluent

Trickling filter plant effluent was taken from the high rate trickling filter operated by the Campus Sewage Treatment Plant. This unit treats about 0.25 MGD of raw sewage. Performance data for this plant are shown in the appendix.

Effluent from this unit was taken from the final settling basin prior to chlorination. This effluent was noticeably different from the two effluents used previously in that it had a gray color and a slight odor. Six runs were made, two under each of the three temperature and pressure conditions. The results of these runs are shown in Tables 11, 12, and 13.

All products had odors as bad or worse than odors of products produced from extended aeration effluent or contact stabilization effluent. As in previous runs, odors tended to increase with increasing operating temperature.

The products produced under high temperature and pressure conditions had a definite hazy appearance. With this one exception, all effluents tested under all operating conditions produced products having a crystal clear appearance.

Results of COD analyses made on these products indicated that organic contamination was significantly worse than had been experienced during previous tests using cleaner feedwaters. Since a significant break-point seemed to have been reached with regards to product water quality, the decision was made to conduct no further single stage tests with poorer quality feedwater.

Activated Carbon Tests

In order to evaluate the effectiveness of activated carbon for removal of any odors remaining after evaporation, a bench scale granular activated carbon column was set up. The carbon selected, Nuchar WV-W, has a large proportion of small pores and is specifically recommended for removal of organics causing taste and odor problems in municipal water.

Since all products produced during the initial single-stage testing had some odor, it was necessary to test examples of products produced under all conditions. Each test consisted of continuously feeding the product water to the carbon column at a rate of about 1 gpm/ft^2 until several column volumes had been flushed through the column. After this flushing, samples were taken of the column feed and product for analysis. The operator would also make observations to determine whether or not any odor remained in the water from the column. In some cases the opinions of several people were secured regarding the odor of the product water.

TABLE 11

INITIAL TESTS, TRICKLING FILTER EFFLUENT,
VACUUM CONDITIONS

	Run No, 1	2
Operating Conditions		
Feed Rate, ml/min	645	640
Evaporation Rate, ml/min	100	97
% Evaporated	15.5	15.2
Hot Water Temp., ^{o}F	142	143
First Stage Temp., ^{o}F	113	114
Feed Temp., ^{o}F	90-95	95-102
Feed, Before Degassing		
Alkalinity, mg/l $CaCO_3$	39	39
pH	6.6	6.6
COD, mg/l	116	116
NH_3-N, mg/l	5.4	5.4
NO_3-N, mg/l	15.2	15.2
Feed, After Degassing		
pH	5.3	5.9
COD, mg/l	101	127
NH_3-N, mg/l	6.0	5.5
NO_3-N, mg/l	13.0	13.4
Product		
pH	5.8	6.6
COD, mg/l	2.0	1.4
NH_3-N, mg/l	0.1	0.6
NO_3-N, mg/l	0.03	0.00
Concentrated Effluent		
pH	6.2	6.5
COD, mg/l	---	123
NH_3-N, mg/l	6.4	6.3
NO_3-N, mg/l	13.4	14.2

TABLE 12

INITIAL TESTS, TRICKLING FILTER EFFLUENT,
ATMOSPHERIC CONDITIONS

	Run No. 1	Run No. 2
Operating Conditions		
Feed Rate, ml/min	645	650
Evaporation Rate, ml/min	172	172
% Evaporated	26.6	26.5
Steam Temp., ^{o}F	228	228
First Stage Temp., ^{o}F	212	212
Feed Temp., ^{o}F	90-102	101-111
Feed, Before Degassing		
Alkalinity, mg/l $CaCO_3$	50	50
pH	7.0	7.0
COD, mg/l	93	93
NH_3-N, mg/l	5.1	5.1
NO_3-N, mg/l	11.4	11.4
Feed, After Degassing		
pH	5.5	6.1
COD, mg/l	93	93
NH_3-N, mg/l	5.9	5.7
NO_3-N, mg/l	10.0	10.8
Product		
pH	5.8	6.5
COD, mg/l	5.8-6.2	6.2-6.6
NH_3-N, mg/l	0.9	1.6
NO_3-N, mg/l	0.06	0.02
Concentrated Effluent		
pH	6.5	7.0
COD, mg/l	105	93
NH_3-N, mg/l	7.2	6.8
NO_3-N, mg/l	12.6	13.2

TABLE 13

INITIAL TESTS, TRICKLING FILTER EFFLUENT,
PRESSURE CONDITIONS

	Run No. 1	Run No. 2
Operating Conditions		
Feed Rate, ml/min	630	630
Evaporation Rate, ml/min	140	165
% Evaporated	22.2	26.2
Steam Temp., ^{o}F	298	300
First Stage Temp., ^{o}F	286	286
Feed Temp., ^{o}F	95-104	103-113
Feed, Before Degassing		
Alkalinity, mg/l $CaCO_3$	65.7	65.7
pH	7.0	7.0
COD, mg/l	90	90
NH_3-N, mg/l	7.5	7.5
NO_3-N, mg/l	16	16
Feed, After Degassing		
pH	5.5	6.1
COD, mg/l	84	80
NH_3-N, mg/l	8.0	8.8
NO_3-N, mg/l	9.4	9.4
Product		
pH	6.0	6.3
COD, mg/l	18.5-20.9	19.6-22.6
NH_3-N, mg/l	2.4	3.7
NO_3-N, mg/l	0.28	0.17
Concentrated Effluent		
pH	6.5	6.6
COD, mg/l	75	76
NH_3-N, mg/l	10.0	10.3
NO_3-N, mg/l	11.6	11.8

Twenty runs were made using the activated carbon column during the initial testing phase. Six were made using products from extended aeration effluent ten using products from contact stabilization effluent, and four using products from trickling filter effluent. In all cases the odors were either eliminated or reduced to such a level that they could be detected by only a minority of the odor test panel. The worst sample tested was a product produced under high temperature and pressure conditions using trickling filter effluent. After activated carbon treatment only one out of five members of the odor test panel detected any odor at all. COD values before and after carbon treatment for this sample were 20.7 and 3.4 mg/1.

The average COD values in and out of the carbon column for the twenty runs were 4.4 and 1.35 mg/1. Omitting the trickling filter product noted above, these averages drop to 3.6 and 1.25 mg/1.

Ammonia Ion Exchange Tests

A bench scale ion exchange column was used to evaluate the feasibility of removal of ammonia from the product water. This column contained clinoptilolite, a natural exchange material selective for ammonium ion. In each test, the column was operated continuously until the column had been thoroughly flushed and then samples were taken of the column feed and product. Surface loading rates of 1.2-1.4 gpm/ft^2 were used. A total of six tests was made using feedwater containing from 2.5-5.0 mg/1 NH_3-N. Ammonia in the column effluent ranged from 0.12-0.32 mg/1 NH_3-N. In all cases ammonia removal exceeded 90% and averaged 94.4%.

Aeration Tests

Probably the most economical means of removing odors from the evaporator product water would be aeration. However, the effectiveness of this method was not known. To evaluate aeration for odor removal, a 13-liter sample of product water was aerated in a 5-gal. carboy using a diffuser stone to disperse the air. The product water used was produced from extended aeration plant effluent under atmospheric pressure and initially had only a mild odor. The sample was aerated up to 1.5 ft^3/gal. This air volume exceeds by an order of magnitude the amount normally used in aeration for odor removal.[41] Little, if any change in odor could be detected in the water.

Because of the poor results from this test, no further evaluations of aeration for odor removal were made.

Product Water Conductivity

Conductivity measurements were made on products produced under atmospheric and pressure conditions using extended aeration effluent and contact stabilization effluent as feedwater. It was found that conductivity correlated closely with the amount of ammonia in the product for products with less than 1.5 mg/1 NH_3-N. Data for 31 product samples are shown in Figure 9.

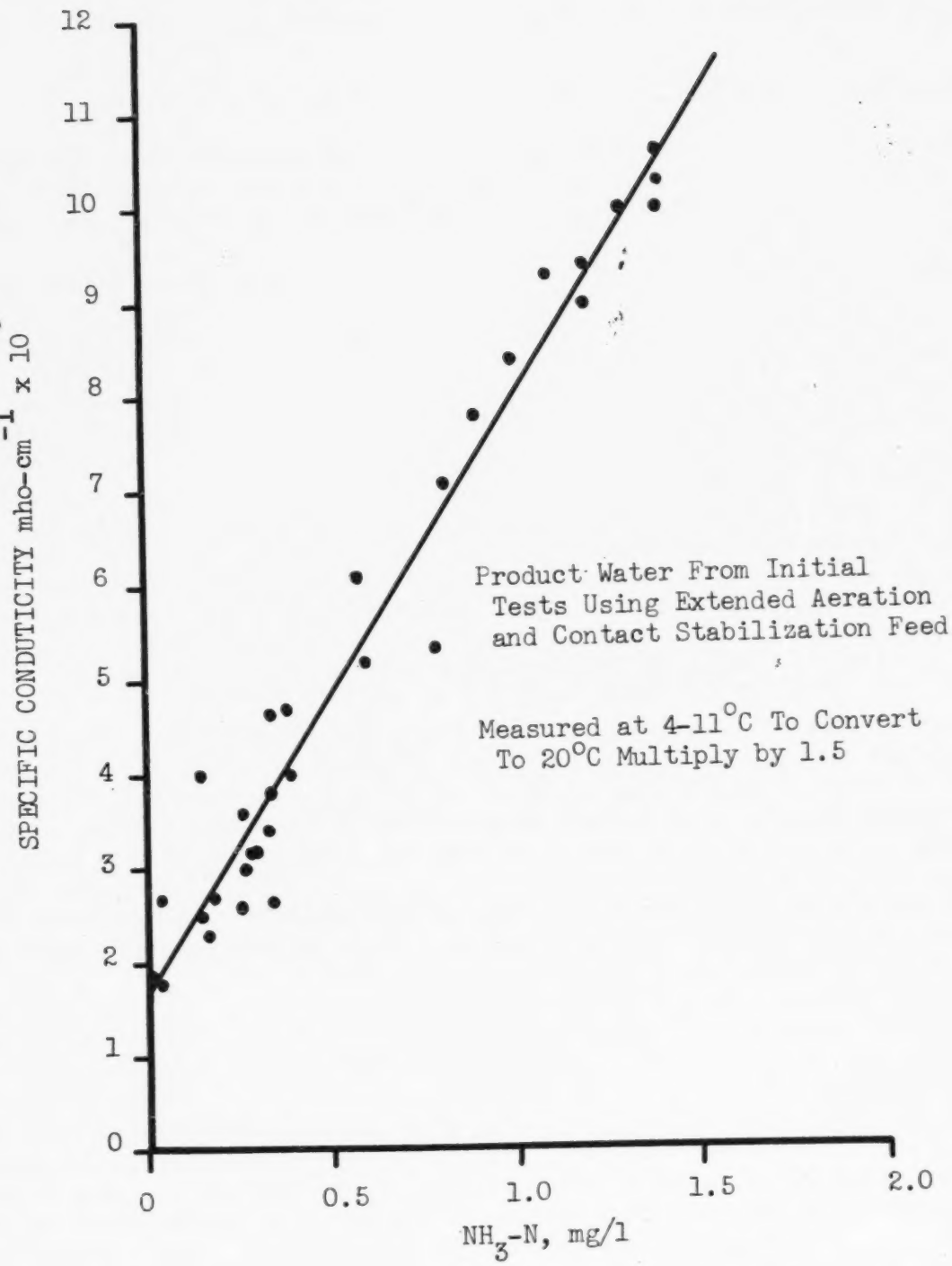

Figure 9. Conductivity vs. Ammonia Concentration

A least squares regression line for these data indicates a conductivity of 1.7×10^{-6} mho/cm at 7°C for zero ammonia. Converting this to 20°C gives a conductivity of about 2.5×10^{-6} mho/cm. By comparison, distilled water in equilibrium with air has a conductivity of about 1×10^{-6} mho/cm and a 1.0 mg/l solution of KCl has a conductivity of about 2×10^{-6} mho/cm.[40]

Scaling During Check-Out and Initial Testing

Although there was no planned program of scale evaluation during the check-out and initial testing phase, the evaporator was mechanically cleaned several times and a few measurements made on the material removed.

During the initial check-out of the system, the evaporator was operated as a single-effect unit (3rd stage only), intermittently, for a total of about 30 hours using tap water from the water treatment plant, Gainesville, Florida. Typical analysis of this water is shown in Table 14. The upper end of the evaporator was inspected by removing the vapor-liquid separator and a light tan scale was observed. The evaporator was not cleaned at this time and was operated, again on an intermittent basis, for an additional 40 hours using tap water with the pH adjusted down to 4.1-6.4 with sulfuric acid. During this period several inspections indicated that the scale was still present but did not seem to be getting any worse. At the end of this period (70 hours total operating time) the evaporator tube was mechanically cleaned with a power driven wire brush and the scale retained. A total of 16.5 grams of dry scale was removed from the evaporator. Approximate analyses on this scale indicated that it was largely calcium carbonate.

Shortly after this cleaning, the degassing equipment was installed and all feed was degassed prior to being put into the evaporator. The evaporator was cleaned a second time after about 65 hours of operation using degassed extended aeration and contact stabilization effluents. This time only 0.92 grams of scale were removed. No analyses were made on this scale.

The evaporator was cleaned a third time after the completion of the initial single stage tests. The scale removed was inadvertently discarded before it could be analyzed.

Long Term Runs

Based on the results of the single stage tests, it appeared that no combination of effluent feedwater and evaporator operating conditions would give a product acceptable for municipal reuse without further treatment. Likewise, all product water appeared amenable to treatment by activated carbon to an aesthetically acceptable quality for reuse. Since it would be economically desirable to operate a full scale unit at as high a temperature and pressure as possible, it was decided to carry out the long term tests under the high temperature and pressure conditions used in the initial tests. All three evaporator effects were used.

The major purpose of these tests was to evaluate the scaling potential of the particular feedwater being used. Since the evaporator was operated

TABLE 14

TYPICAL ANALYSIS OF TAP WATER,
GAINESVILLE, FLORIDA

Total Hardness	95 mg/l as $CaCO_3$
Calcium Hardness	48 mg/l as $CaCO_3$
Magnesium Hardness	47 mg/l as $CaCO_3$
Total Alkalinity	73 mg/l as $CaCO_3$
Sulfate	32 mg/l as SO_4
Chloride	20 mg/l as Cl
Total Dissolved Solids	211 mg/l
pH	8.4

as a three-effect unit, it was possible to evaporate a greater percentage of the feedwater and thus observe the scaling that might occur in the latter stages of a full scale unit. After each extended run, the scale in each stage was mechanically removed and analyzed. In addition, the overall heat transfer coefficients for each effect were calculated as a function of time. Thus it was possible to get some indication of scaling during a run without having to shut down the equipment.

In addition to information on scaling, the long term runs gave additional product quality data over a long period of time. Also, quality data were obtained on products from the second and third evaporator effects. Previous single-effect tests had given information only on the first-effect product quality.

Operating and sampling procedures during the long term runs were as follows. Fresh feed for the evaporator was prepared every 8 hours by degassing approximately 100 gal. of fresh effluent. Thus two 55-gal. drums of feed were used every 8 hours. Samples of the effluent being used, before degassing, were taken once every 8 hours and composited for 24 hours. Samples of the degassed feed and concentrated effluent were taken every 4 hours and composited for 12 hours. All the product water from each stage was collected continuously and measured and sampled each hour. The product samples were composited for 12 hours.

First Extended Aeration Effluent Run

The first long-term run was carried out using extended aeration effluent. It was planned to operate with 300°F steam and adjust the third stage pressure to evaporate about 75% of the incoming feed. The evaporator was operated for a total of 323 hours (13 days, 11 hours). This excluded 9 hours downtime to make minor repairs on the equipment. Other than these two brief shutdowns, one to tighten a leaking fitting and another to repair a liquid seal, the run was reasonably routine. Trouble was periodically encountered with maintaining the incoming steam to the first effect at 300°F. It was determined that there were actually two problems. First, the first-effect pressure was too near the supply line steam pressure, and at times there was not enough pressure differential across the steam regulator to maintain reliably the first-effect pressure. Second, the steam trap in the condensate line from the first stage was oversized and allowed a 5-10°F temperature drop every time it opened to discharge condensate. Consequently, on the eighth day of the run the steam temperature to the first effect was lowered to 292°F. This gave better control, but some temperature variation was still caused by the oversized trap. This problem had little effect on the validity of the scaling or product quality results, but it did make it rather difficult to determine individual heat transfer coefficients.

During the run an attempt was made to maintain a temperature drop of 32-34°F from incoming steam to third-stage vapor-liquid separator temperature. At a constant feed rate of 630 ml/min this gave an average product rate of 422 ml/min or 67% of the feed rate. The product rates are shown as a function of time in Figure 10. Problems with steam control caused some variations in product rates. However, there were other variations that could not be

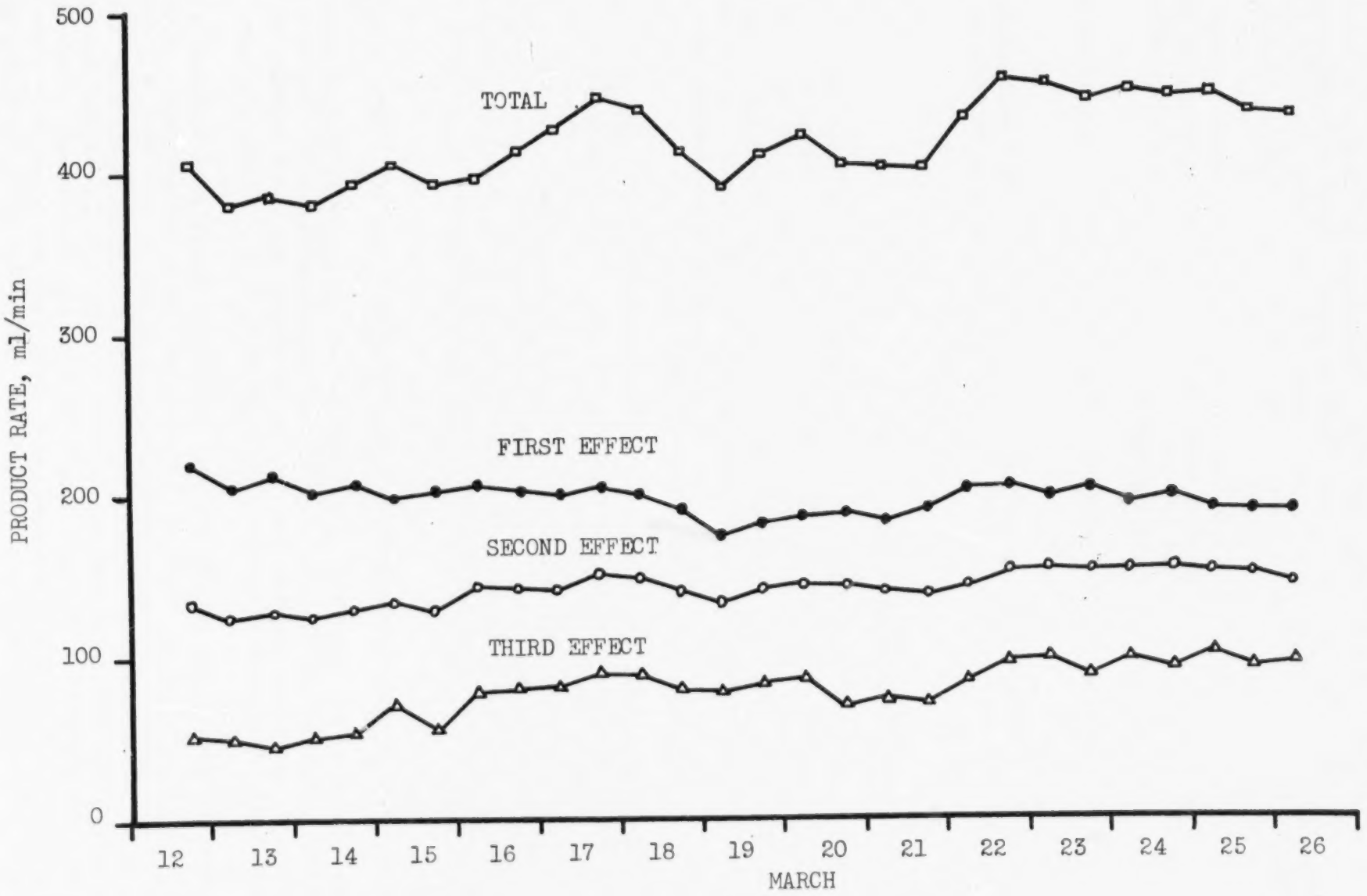

Figure 10. Product Rates, First Extended Aeration Effluent Run

accounted for by this factor. Overall, there was a slight increase in the total product rate over the run. This increase was due to increases in the rates from the second and third effects. These increases apparently were the result of decreased heat losses which resulted when the slightly damp insulation on these effects dried. There was a cycling pattern in the output rate that occurred within the individual effects as well as with the total. It was noted that the product rate occasionally increased markedly immediately following some type of temporary operational upset. Also, there were times that the concentrated effluent contained unusually high levels of dark-colored, suspended solids.

The analytical results for samples taken during this run are shown in Table 15.

After the completion of the run the evaporator tubes were mechanically cleaned and the scale retained. Results of the analysis of this scale are shown in Table 16.

Trickling Filter Effluent Run

Based on the good results of the extended aeration effluent run, it was decided to make an extended run under pressure conditions using trickling filter effluent. Prior to the start of this run the oversized steam trap in the condensate line from the first stage was replaced with a smaller trap. This change made it possible to control the incoming steam temperature closer to 292°F. The operational and sampling procedures were the same as were used during the first extended aeration run.

The run lasted 285 hours (11 days, 21 hours). Total downtime was one hour. The feed rate was held constant at 630 ml/min. For the overall run the total product rate averaged 457 ml/min or 72.6% of the feed rate. Daily averages for the total product rate ranged from 404 to 497 ml/min or 64.2 to 79.0% of the feed rate. The overall trend in total product rate was downward. As in the previous test, cycling in the product output rate and periodic increases in suspended solids in the concentrated effluent were observed.

Analytical results for samples taken during this run are shown in Table 17.

Close control of operating temperatures made it possible to calculate accurate overall heat transfer coefficients for each effect. The 12-hour averages for these results are plotted in Figure 11. In general, the coefficients decreased with time, but tended to show a cyclic pattern.

After the completion of the run the evaporator tubes were mechanically cleaned and the scale retained. As had been experienced before, most of the scale was removed by the power-driven wire brush in less than 5 minutes. However, there was some harder gray scale in the upper portion of the evaporator tube in effects two and three. Most of this scale was removed only after considerable additional scrubbing with the wire brush. This "hard" scale was collected and analyzed separately from the previously removed "soft" scale. Results of the analysis of these scales are shown in Table 18.

TABLE 15

ANALYTICAL RESULTS - FIRST EXTENDED AERATION EFFLUENT RUN

		COD	NH_3-N	NO_3-N	Total Dissolved Solids [a]	Suspended Solids [a]	Total Hardness [b]	pH
Feed								
Before Degassing	Avg.	46.5[c]	0.37	17	404	20	---	6.6
	Range	35.4-63.8	0.28-0.59	12-21	398-410	15-24	---	6.3-7.0
After Degassing	Avg.	39.0	0.36	18	388	16	95.3	5.8
	Range	28.3-49.4	0.20-0.62	12-24	354-428	10-20	93.6-97.8	5.5-6.2
Concentrated Effluent	Avg.	105	1.4	55	1,084	56	317	6.2
	Range	72-211	0.9-2.2	42-70	938-1,222	34-101	---	5.8-6.6
Products								
#1	Avg.	5.9	0.12	0.14	---	---	---	---
	Range	1.2-14.7	0.05-0.19	0.00-0.23	---	---	---	---
#2	Avg.	5.0	0.13	0.10	---	---	---	---
	Rage	1.8-8.6	0.08-0.21	0.00-0.17	---	---	---	---
#3	Avg	4.3	0.10	0.15	---	---	---	---
	Range	0.8-9.5	0.03-0.22	0.00-0.55	---	---	---	---

[a] Spot checks on less than 5 samples
[b] As $CaCO_3$; spot check on 8 feed samples and 1 concentrated effluent sample
[c] All concentrations in mg/l except pH

TABLE 16

SCALE FROM FIRST EXTENDED AERATION EFFLUENT RUN[a]

	Stage No. 1	2	3
Total Weight, gms	0.83	1.04	3.65
% Benzene Soluble[b]	2.5	3.8	4.3
% Organic[c]	39.7	53.1	35.2
Analysis of Fired Residue			
% Soluble in 0.5N HCl	78.1	65.8	91.0[d]
% Calcium as Ca[e]	14.3	11.3	12.4
% Magnesium as Mg[e]	1.4	1.6	0.7
% Sodium as Na[e]	1.0	0.7	1.3
% Iron as Fe[e]	13.5	1.8	8.0
% Sulfate as SO_4[e]	0.0	5.1	17.7
% Phosphate as PO_4[e]	18.5	26.5	13.5

[a] Samples of 0.15 to 0.25 grams were stirred in one liter of 0.5 N HCl for several hours

[b] Soxhlet Extraction

[c] Benzene soluble + loss after firing at 600°C

[d] More severe conditions used to dissolve this sample

[e] Expressed as percentage of soluble inorganic scale

TABLE 17

ANALYTICAL RESULTS - TRICKLING FILTER EFFLUENT RUN

		COD	NH_3-N	NO_3-N	Total Solids[a]	pH
Feed						
Before Degassing	Avg.	112[b]	12.1	6.3	---	7.1
	Range	84-133	2.7-17	0.7-11	---	6.7-7.4
After Degassing	Avg.	109	14	7.7	456	5.7
	Range	77-140	7.6-20	4.4-12	365-562	4.8-6.5
Concentrated Effluent	Avg.	286	40	35	1,635	5.8
	Range	202-367	19-66	15-64	1,405-2,302	5.0-6.4
Products						
#1	Avg.	22.2	2.9	0.08	---	---
	Range	7.6-35.5	1.1-3.8	0.04-0.19	---	---
#2	Avg.	12.0	2.4	0.06	---	---
	Range	7.4-23.6	1.0-3.1	0.03-0.11	---	---
#3	Avg.	9.4	2.7	0.10	---	---
	Range	4.0-15.4	1.2-3.9	0.04-0.14	---	---

[a] Spot checks on 5 samples each
[b] All concentration in mg/l except pH

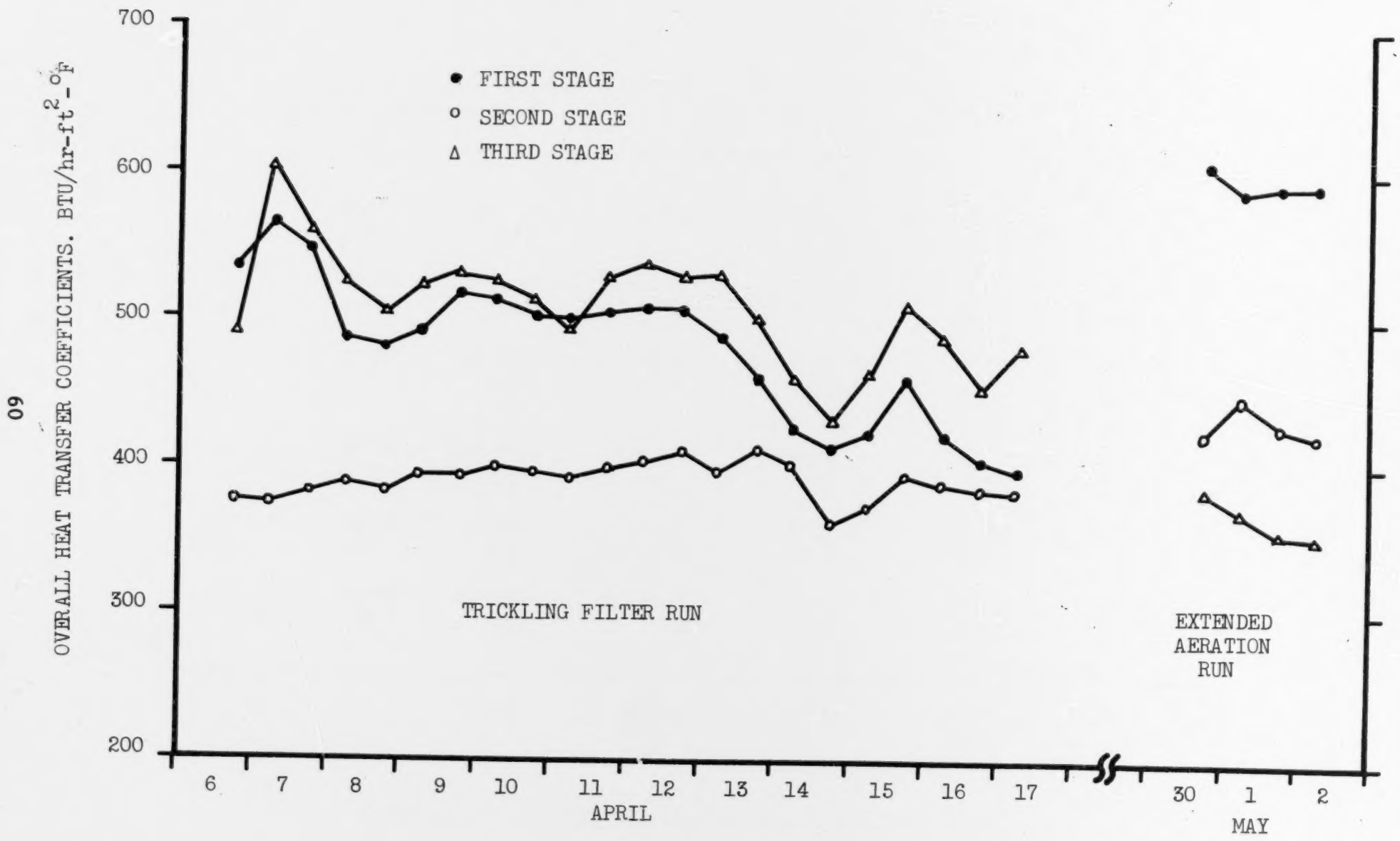

Figure 11. Overall Heat Transfer Coefficients

TABLE 18

SCALE FROM TRICKLING FILTER EFFLUENT RUN[a]

	"Soft" Scale, Stage			"Hard" Scale, Stage	
	1	2	3	2	3
Total Weight	5.01	3.04	6.88	0.25	0.94
% Organic[b]	52.4	63.5	54.4	35.8	26.9
Analysis of Fired Residue					
% Soluble in 0.5N HCl	79.5	68.1	80.0	96.7	93.3
% Calcium as Ca[c]	7.7	15.5	20.9	21.8	23.0
% Magnesium as Mg[c]	0.8	1.4	0.7	0.3	0.1
% Sodium as Na[c]	2.6	3.4	0.8	0.8	0.6
% Iron as Fe[c]	19.4	3.8	4.4	0.3	0.0
% Sulfate as SO_4[c]	7.3	20.0	38.2	54.8	51.5
% Phosphate as PO_4[c]	18.4	17.1	8.5	13.3	4.2

[a] Samples of 0.15 to 0.25 grams were stirred in one liter of 0.5 N HCl for several hours.

[b] Loss after firing at 600°C

[c] Expressed as percentage of soluble inorganic scale

TABLE 19

ACTIVATED CARBON TREATMENT OF PRODUCTS FROM TRICKLING FILTER EFFLUENT RUN

	COD, mg/l	
	Before Carbon Treatment	After Carbon Treatment
Product from Effect #1	20.1	4.9
Product from Effect #2	11.6	3.3
Product from Effect #3	10.6	0.1

To confirm the initial single stage test results with regard to odor removal, samples of each product were treated in the activated carbon column. The same column set-up used previously was used in these tests. The column loading rate for all three products was 1.0 gpm/ft. The odor was completely removed from all three products. COD results are shown in Table 19.

Second Extended Aeration Effluent Run

A reasonable explanation for the cycling in product output and heat transfer coefficients observed during the first two long term runs seemed to be that relatively soft scale was alternately building up and then flaking off the evaporator tube walls. The periodic increases in suspended solids in the concentrated effluent were further evidence that this was occurring. To test further this theory, a second run using extended aeration effluent was made. This test was limited to 5 days. If, after 5 days, quantities of scale were found well out of proportion of what would be expected based on the previous 14-day test, this would further confirm that scale was alternately building up and flaking off.

The operational and sampling procedures used in this test were identical to those used in the first two runs. However, from the beginning of the run there was an abnormally high product rate from the third effect. The liquid seal between the second and third effects appeared to be leaking, allowing a small amount of vapor from the second vapor-liquid separator to enter the bottom of the third effect. It was felt that this would have little if any effect on the scaling rate and the decision was made not to shut down for repairs. However, the problem appeared to worsen with time, and on the third day of the run the unit was shut down for 5 hours and the liquid seal repaired. Other than the problem with this seal, the run was rather routine. The evaporator was operated for a total of 113 hours (4 days, 17 hours). Total downtime was 5 hours.

Analytical results for samples taken during this run are shown in Table 20.

Heat transfer coefficients for the three effects were calculated for the latter part of the run after the liquid seal was repaired. These results, calculated as 12-hour averages, are shown in Figure 11.

During the latter part of the run, three complete sets of samples were taken for bacteriological analyses. Results of these tests are shown in Table 21. In addition to the coliform tests, a test designed to check bacterial sterility was run on all samples. In this test the sample is incubated at 35°C for seven days in nutrient broth. This test revealed that all product samples from the first and second effects were sterile. All other samples showed some biological activity.

After the run was completed the evaporator was mechanically cleaned and the scale retained. Results of analysis of this scale are shown in Table 22.

TABLE 20

ANALYTICAL RESULTS - SECOND EXTENDED AERATION EFFLUENT RUN

		COD	NH_3-N	NO_3-N	Total Solids[a]	pH
Feed						
Before Degassing	Avg.	74[b]	0.39	14	---	6.8
	Range	38-97	0.25-0.52	14-15	---	6.7-7.0
After Degassing	Avg.	35	0.37	14	335	5.7
	Range	29-43	0.28-0.51	14-16	328-341	5.5-6.0
Concentrated Effluent	Avg.	136	2.3	79	1,643	5.8
	Range	76-203	1.6-3.7	62-98	1,265-2,075	5.5-6.7
Products						
#1	Avg.	7.4	0.08	0.09	---	---
	Range	3.4-13.2	0.00-0.11	0.05-0.13	---	---
#2	Avg.	4.6	0.12	0.13	---	---
	Range	1.4-8.2	0.06-0.17	0.04-0.26	---	---
#3	Avg.	4.0	0.08	0.00	---	---
	Range	0.9-7.5	0.02-0.11	0.00	---	---

[a] Spot check of 5 samples each
[b] All concentration in mg/l except pH

TABLE 21

BACTERIOLOGICAL TESTS - SECOND EXTENDED AERATION
EFFLUENT RUN

Sample	Time	Date	Total Coliform MPN/100 ml	Fecal Coliform MPN/100 ml
Extended Aeration Plant Effluent	2400	4/30/70	1,300,000	790,000
Feed	2400	4/30/70	< 2.0	< 2.0
Product No. 1	2330	4/30/70	< 2.0	< 2.0
Product No. 2	2330	4/30/70	< 2.0	< 2.0
Product No. 3	2330	4/30/70	< 2.0	< 2.0
Concentrated Effluent	2400	4/30/70	< 2.0	< 2.0
Extended Aeration Plant Effluent	0400	5/1/70	490,000	330,000
Feed	0400	5/1/70	8.0	2.0
Product No. 1	0330	5/1/70	< 2.0	< 2.0
Product No. 2	0330	5/1/70	< 2.0	< 2.0
Product No.3	0330	5/1/70	< 2.0	< 2.0
Concentrated Effluent	0400	5/1/70	< 2.0	< 2.0
Extended Aeration Plant Effluent	0800	5/1/70	490,000	109,000
Feed	0800	5/1/70	< 2.0	< 2.0
Product No. 1	0730	5/1/70	< 2.0	< 2.0
Product No. 2	0730	5/1/70	< 2.0	< 2.0
Product No. 3	0730	5/1/70	< 2.0	< 2.0
Concentrated Effluent	0800	5/1/70	< 2.0	< 2.0

TABLE 22

SCALE FROM SECOND EXTENDED AERATION EFFLUENT RUN[a]

	Stage No. 1	2	3
Total Weight, gms	0.92	2.54	1.89
% Organic[b]	30.7	36.4	28.4
Analysis of Fired Residue			
% Soluble in 0.5N HCl	70.2	87.7	87.9
% Calcium as Ca[c]	13.0	19.2	24.4
% Magnesium as Mg[c]	0.8	0.8	0.6
% Sodium as Na[c]	0.6	0.7	0.9
% Iron as Fe[c]	14.2	2.0	2.0
% Sulfate as SO_4[c]	0.9	30.1	32.4
% Phosphate as PO_4[c]	17.8	19.2	11.9

[a] Samples of 0.15 to 0.25 grams were stirred in one liter of 0.5 N HCl for several hours

[b] Loss after firing at 600°C

[c] Expressed as percentage of soluble inorganic scale

Three Effect Ammonia Distribution Tests

Some indication of ammonia carry-over as a function of feed pH and evaporator operating conditions was gained in the single-effect tests. However, additional information was needed regarding the distribution of ammonia between product and concentrated effluent in subsequent evaporator effects.

To obtain this information, a series of tests was made utilizing all three evaporator effects. Degassed contact stabilization effluent was used as the feedwater. The pH of the feed was varied from 6.0 to 8.7 and the evaporator was operated at least 4 hours under each pH condition. Product samples were taken from the last 2-hour's production and composited. Feed and concentrated effluent samples were taken at the end of each run. Feed rate was held constant at 630 ml/min and the total evaporation rate varied between 447 and 545 ml/min or 71.0-86.5% of the feed rate.

The analytical results for samples taken during these runs are shown in Table 23.

TABLE 23

ANALYTICAL RESULTS - THREE EFFECT AMMONIA DISTRIBUTION TESTS

		Test No. 1	2	3	4	5	6
Feed After Degassing	pH	6.0	6.7	7.3	7.6	8.0	8.7
	NH_3-N, mg/l	8.3	8.0	8.2	2.8	9.2	2.3
	COD, mg/l	35	49	53	24	26	23
Concentrated Effluent	pH	5.8	--	--	7.3	5.9	6.9
	NH_3-N, mg/l	35	18	16	2.8	21	3.4
	COD, mg/l	147	184	160	57	137	65
Products #1	NH_3-N, mg/l	2.7	5.0	4.9	1.7	6.3	1.4
	COD, mg/l	16.3	4.8	5.6	1.0	4.7	4.4
#2	NH_3-N, mg/l	1.9	2.9	2.9	1.0	3.5	1.0
	COD, mg/l	3.3	8.0	0.6	0.3	3.2	1.6
#3	NH_3-N, mg/l	2.2	2.2	2.1	0.8	2.6	0.8
	COD, mg/l	4.5	4.5	1.8	2.6	0.9	0.4

DISCUSSION OF RESULTS

Product Quality

The first objective of this work was to define the relationships between product water quality, feedwater quality, and evaporation conditions. The second objective involved defining the relationships between the above variables and post evaporation polishing. The three primary parameters used to measure product quality were ammonia content, COD, and odor. Of secondary importance in measuring product quality were conductivity and nitrate content. Product quality with regard to each of these parameters is discussed separately below.

Ammonia Content

Ammonia is an undesirable constituent in the product water because it consumes more than six times its weight of chlorine and produces chloramines which have much lower disinfecting efficiency than free chlorine.[16] Previous laboratory investigations had indicated that only by going to extremely acid or basic conditions could ammonia distribution between the evaporator feedwater and the product water be controlled.[7] In the work reported here, feedwater pH values between 5.1 and 8.7 were used in both single-effect and three-effect tests. The test results show that for a given pH value, the fraction of ammonia carrying over to the product increases as operating temperature and pressure increase. This is consistent with the theoretical distribution of ammonia and ammonium ion as a function of temperature.

As can be seen from the results of the single-stage tests using contact stabilization effluent and trickling filter effluent, under vacuum conditions, ammonia in the product was less than 1 mg/l for feed pH values up to around 6.5. Under atmospheric conditions, feed pH had to be less than 6 to control ammonia in the product to less than 1 mg/l. It should be noted that these results are for the first-effect product only. In all the single-effect tests under vacuum and atmospheric conditions using contact stabilization or trickling filter effluent as feedwater, the pH of the concentrated effluent was significantly higher than the feedwater pH. This would mean that under the multieffect conditions that would be used in full scale applications, ammonia could be expected to contaminate the product water in the later stages.

Under pressure conditions, there was more than 1 mg/l ammonia in all products from contact stabilization or trickling filter feedwater for pH values down to 5.1. Tests made while operating the evaporator as a three-effect unit showed that this contamination persisted in the second and third effects. Raising the feedwater pH as high as 8.7 did not eliminate contamination in the second and third effects.

There appear to be no easy ways of controlling ammonia contamination of the product water short of removing the ammonia from the feedwater. This could be accomplished by using a highly nitrified effluent as feedwater.

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The tests reported here and the work of other investigators[42,43] show that ammonia could be removed from the product water by ion exchange using clinoptilolite. If the product water were essentially free of cations other than ammonium, ordinary cation resins could also be used for this purpose.

Chemical Oxygen Demand

The COD test was used as an indicator of the organic content of the product water. For the products from extended aeration effluent and contact stabilization effluent, the COD was generally less than 5 mg/1 for all evaporator operating conditions. However, when trickling filter effluent was used, the product COD increased significantly as operating temperature and pressure increased. Thus it appears, that because of problems with organic carry-over to the product, trickling filter effluent is a significantly poorer feedwater than either of the other two effluents tested.

The COD of products from all three effluents at all operating conditions was substantially reduced by treatment with activated carbon. However, for product that was produced under pressure conditions from trickling filter effluent, the final COD after activated carbon treatment was roughly equivalent to the COD of the other products before carbon treatment.

Odor

Odor in the product water would be a particularly critical factor in any full scale application of direct wastewater reuse because of the obvious psychological barriers. In these tests, no combination of wastewater feed and evaporation conditions produced completely odor free water. Product odor tended to increase in intensity and disagreeableness as evaporator temperature and pressure increased. Also, the more completely treated effluents tended to produce less odorous products.

Activated carbon was effective in removing the odor from all products. Aeration was ineffective in removing product odors.

There was speculation regarding the effectiveness of the degassing operation in eliminating odorous compounds. Based on the relatively mild odor of the degasser condensate and the odor of products from the degassed feedwater, the degassing operation seems to have little effect on product odor.

Conductivity

Conductivity was measured on some of the products from the initial testing. As shown in Figure 9, conductivity was generally related to the ammonia content of the water. However, extrapolation to zero ammonia shows the water was of extremely high quality with regard to inorganic salts.

Nitrate

Nitrate in the product water was used primarily as an indicator of physical carry-over to the product, i.e. incomplete separation of vapor from liquid

in the vapor-liquid separator. In practically all cases, nitrate test results showed that physical carry-over was less than 0.5% of the incoming feedwater and in many cases much less than this.

Since it can be shown that product waters were not significantly contaminated by physical carry-over, the organic contamination that did occur resulted from evaporation of volatile organics and subsequent condensation.

Bacteriological Tests

The bacteriological tests generally confirm that water safe for human consumption from the standpoint of pathogen content can be produced by evaporation. In all cases, product water from effects one and two was sterile. Considering this, it appears likely that the non-sterility of the product water from effect three was due to contamination in the final condenser and receiver system. In any case, the product water would be further treated by chlorination before use.

Evaporator Scaling

The third objective of this work was to define the relationships between feedwater quality, evaporator operating conditions, and scaling or fouling of the evaporator heat transfer surfaces.

The first indication of the potential seriousness of calcium carbonate scaling came during the initial check-out of the evaporator. Tap water that had not been degassed proved to be extremely scale forming. The 16.5 grams of scale removed from the third effect after the first 70 hours of operation was more than the total amount of scale removed from that effect during the entire remainder of the testing program. Thus it appears that degassing of evaporator feedwater to remove carbonates is essential.

The results of the three long term runs indicate that scaling from trickling filter effluent is markedly worse than scaling from extended aeration effluent. These are discussed separately below.

Scaling by Extended Aeration Effluent

Two long term runs, one lasting about 13 days and the other lasting about 5 days, were made using extended aeration effluent. The purpose of the second run was to further investigate the possibility that scale was alternately building up on and flaking off the evaporator heat transfer surfaces. The results of the second run seem to confirm this. The amount of scale in effects one and two after 5 days was more than had been removed from these two effects at the end of the 13 day run. These results, the cycling effects observed in product output rates and heat transfer coefficients, the periodic increase in suspended solids in the concentrated effluent, and the occasional increases in product rates observed after operational upsets, all indicate that the scale was lightly adhering and would periodically flake off. Thus it appears that scaling problems with extended aeration effluent would be minor.

Analyses of the scale showed it to be about one-third to one-half organic. The inorganic portion contained significant quantities of calcium, iron, sulfate, and phosphate. It is probable that the iron came from the mixing tank and associated piping when the effluent pH was lowered for degassing. With proper materials of construction in this part of the process, iron should not be a problem. Calcium sulfate scale was formed only in the second and third effects. This reflects the increase in concentration caused by evaporation. It should be noted that, during the long term tests, conditions in the latter two effects were more severe than would occur in full scale application of this process. In a full scale plant, the evaporation of the same fraction of the incoming feed would occur over a larger number of effects and hence the temperature of the more concentrated liquid would be appreciably lower than the 260-270°F used in these tests. At the lower temperature, calcium sulfate would be less likely to form scale.

Regarding calcium sulfate scale, consideration should be given to the fact that the pH adjustment prior to degassing was made with sulfuric acid. Acid additions during the two runs ranged from 11 to 35 mg/l SO_4. If higher evaporation temperatures are to be attempted, this amount of extra sulfate could become critical.

Significant amounts of phosphate were found in the scale from all three effects. Because phosphate compounds tend to form loosely adhering sludges rather than hard scales, a high phosphate level should be desirable.

Scaling by Trickling Filter Effluent

During the 12-day run using trickling filter effluent, the scale formed was more undesirable, both with regards to quantity and quality. Significantly more scale formed in all three effects, and the "hard" scale formed in effects two and three was very difficult to remove.

Analyses of the "soft" scale revealed that it was more than half organic, whereas the extended aeration effluent scale was generally less than half organic. The inorganic portion of the "soft" scale was very similar in composition to the extended aeration effluent scale. The "hard" scale removed from effects two and three contained a much greater inorganic fraction that appeared to be largely calcium sulfate. This finding was consistent with the physical characteristics of the material. The greater problem with calcium sulfate experienced with this feedwater may have been caused by the much larger amount of sulfuric acid that was required to lower the pH to 4.9 prior to degassing than was used in the extended aeration effluent runs. An average of 103 mg/l SO_4 was added to the trickling filter effluent prior to degassing.

Heat Transfer Coefficients

The heat transfer data in Figure 11 show that during the trickling filter effluent run, the heat transfer coefficients in effects one and three generally declined and were higher than the rather stable coefficient in effect two. During the extended aeration run, the coefficients decreased

from effect one to effect three. It is likely that during the trickling filter run the coefficients for effect three were high because of a small leak of steam from the second effect through the liquid seal. It was immediately after this run that this leak really became noticeable and was repaired. After the repair, the coefficient distribution seen in the second extended aeration run resulted.

Theoretically, there should be a slight drop in heat transfer coefficient from one effect to the next due to decreasing temperature. However, the differences in coefficients seen during the extended aeration run seem to be unrealistically high. It is probable that the coefficients calculated for effects two and three are lower than the actual values. This is because the steam side temperature in these effects was measured in the vapor-liquid separator of the previous effect. Any temperature drop that occurred between the vapor-liquid separator and the steam jacket of the following effect was not accounted for in the calculations. The calculation for the coefficient in the final effect involved a correction for heating the feed to the boiling point (see Appendix II). Although the correction was made as accurately as possible, it may have contributed a small error to the coefficients. Hence, the coefficients shown in Figure 11 are of more value as relative indicators of gain or loss of heat transfer efficiency than as absolute measures of heat transfer coefficients.

APPENDIX I.

TEST RESULTS - CAMPUS SEWAGE TREATMENT PLANT

Effluent from the High Rate Trickling Filter and the Contact Stabilization unit of the Campus Sewage Treatment Plant was used in this study. Shown in Tables 24 and 25 are results of tests made in the campus plant before, during, and after the time effluent from this plant was being used.

TABLE 24

CAMPUS SEWAGE TREATMENT PLANT TEST RESULTS,
HIGH RATE TRICKLING FILTER

	Alkalinity as $CaCO_3$		Chloride as Cl		BOD		Suspended Solids	
Date	In	Out	In	Out	In	Out	In	Out
2-18-70	--	--	--	--	184	47	--	--
2-19-70	154	82	142	140	--	--	--	--
2-25-70	--	--	--	--	212	38	168	63
2-26-70	160	84	146	144	--	--	--	--
3-4-70	--	--	--	--	179	36	166	52
3-5-70	164	76	146	126	--	--	--	--
3-11-70	--	--	--	--	208	51	181	59
3-12-70	148	74	210	192	--	--	--	--
4-1-70	--	--	--	--	162	37	136	18
4-2-70	148	88	188	168	--	--	--	--
4-8-70	--	--	--	--	198	36	182	56
4-9-70	142	72	134	144	--	--	--	--
4-15-70	--	--	--	--	190	40	125	44
4-22-70	--	--	--	--	--	--	177	50
4-23-70	150	84	132	140	--	--	--	--
Average	152	80	157	151	191	41	162	49

All quantities expressed as mg/l
Initial tests were made during the period 2-27-70 to 3-3-70.
The long term run was made 4-6-70 to 4-17-70.

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TABLE 25

CAMPUS SEWAGE TREATMENT PLANT TEST RESULTS,
CONTACT STABILIZATION PLANT

Date	Alkalinity as $CaCO_3$		Chloride as Cl		BOD		Suspended Solids	
	In	Out	In	Out	In	Out	In	Out
2-4-70	--	--	--	--	202	17.7	152	15.0
2-5-70	120	118	128	120	--	--	--	--
2-11-70	--	--	--	--	168	10.4	179	15.7
2-12-70	108	112	132	140	--	--	--	--
2-18-70	--	--	--	--	185	11.7	--	--
2-19-70	136	114	134	128	--	--	--	--
2-25-70	--	--	--	--	184	9.5	139	14.5
2-26-70	140	120	136	126	--	--	--	--
3-4-70	--	--	--	--	187	7.1	165	8.7
3-5-70	128	116	142	120	--	--	--	--
3-11-70	--	--	--	--	209	6.8	165	11.2
3-12-70	104	106	162	138	--	--	--	--
Average	123	114	139	129	189	10.5	160	13.0

All quantities expressed as mg/l
The initial tests were made during the period 2-11-70 to 3-5-70.

APPENDIX II.

CALCULATION OF HEAT TRANSFER COEFFICIENTS

In calculating overall heat transfer coefficients for the evaporator the following procedures were used.

For the first effect, where the feed entered considerably below its boiling temperature, two coefficients were involved. The first coefficient was for heat transfer from condensing steam to water. The second coefficient was for heat transfer from condensing steam to boiling water and theoretically should be much larger than the first coefficient.

To solve for the two coefficients it was necessary to operate the evaporator at two slightly different steam temperatures. Assuming that the coefficients were constant over the temperature range used, the two coefficients were calculated as follows.

	Condition 1	Condition 2
Steam Temperature	T_{S1}	T_{S2}
Boiling Temperature	T_{B1}	T_{B2}
Feed Temperature	T_{F1}	T_{F2}
Feed Rate, gpm	F_1	F_2
Product Rate, gpm	P_1	P_2

Heat transferred in the heating the feed to boiling was designated Q_h

$$Q_h = (F)\ (8.34)\ (60)\ (T_B - T_F) \quad \text{BTU/hr}$$

If the total area for heat transfer was A and the area used for heating the liquid to its boiling point was A_h

$$Q_h = U_h A_h \overline{\Delta T_L} \qquad (8)$$

where U_h = overall heat transfer coefficient for heating

$\overline{\Delta T_L}$ = log mean temperature difference

$$= \frac{(T_S - T_F) - (T_S - T_B)}{\ln \frac{(T_S - T_F)}{(T_S - T_B)}} \qquad (9)$$

Heat transferred in boiling the liquid was designated Q_B

$$Q_B = (P)(8.34)(60)(h) \quad BTU/hr \qquad (10)$$

where h = heat of vaporization at T_B

Then

$$Q_B = U_B(A-A_h)(\Delta T) \qquad (11)$$

where U_B = overall heat transfer coefficient for boiling

$$\Delta T = T_S - T_B \qquad (12)$$

Assuming that the coefficients are constant over the temperature range used for condition one and two, equations (8) and (11) can be expressed as follows:

$$U_h = \frac{Q_{h1}}{(A_{h1})(\overline{\Delta T_{L1}})} = \frac{Q_{h2}}{(A_{h2})(\overline{\Delta T_{L2}})} \qquad (13)$$

$$U_B = \frac{Q_{B1}}{(A-A_{h1})(\Delta T_1)} = \frac{Q_{B2}}{(A-A_{h2})(\Delta T_2)} \qquad (14)$$

Since for the right hand portions of equations (13) and (14) only A_{h1} and A_{h2} are unknown, these two equations can be used to solve for A_{h1} and A_{h2} and hence for U_h and U_B.

This was done for several sets of conditions, and it was found that U_B was consistently about three times as great as U_h

Using this ratio of U_h to U_B, the coefficients can be solved for at any single condition as follows:

$$U_h = \frac{U_B}{3} = \frac{Q_h}{(A_h)\ (\overline{\Delta T_L})} \tag{15}$$

$$U_B = \frac{Q_B}{(A-A_h)(\Delta T)} \tag{16}$$

Substituting and simplifying

$$A_h = \frac{(3)(Q_h)(\Delta T)(A)}{(Q_B)(\overline{\Delta T_L}) + (3)\ Q_h\ (\Delta T)} \tag{17}$$

Substituting A_h in equation (15) or (16) gives U_B.

For the second and third effects where the liquid enters at very near its boiling point, the assumption was made that only heat transfer from condensing steam to boiling water was occurring and

$$U_B = \frac{Q_B}{(A)(\Delta T)} \tag{18}$$

To save time, all calculations were made using this procedure on an IBM 360 computer.

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27 Abstract

The technical feasibility of evaporation of municipal sewage treatment plant effluent for the purpose of water reuse was investigated. The equipment used was a three-effect long tube vertical (LTV) evaporator. The objectives of the research were to determine the effects of feedwater quality, and evaporation conditions on product water quality, post evaporation polishing, and evaporator tube scaling. Feedwaters tested in the evaporator included extended aeration effluent, high rate trickling filter effluent, and contact stabilization effluent. All feedwater was acidified and vacuum degassed.

Results showed that an odor free product could not be produced from any of the three feedwaters under any operating condition from 112°F to 290 F. The odor intensity increased as the evaporator operating temperature and/or the chemical oxygen demand of the feedwater increased. In all cases, post treatment with activated carbon removed all odors. Aeration would not remove all odors. Product produced under pressure conditions using trickling filter feed contained significantly more organic contamination than any other products. Product contamination by ammonia could not be controlled by adjusting the pH of the feed in the range 5.1 to 8.7. However, ammonia in the product water was removed by ion exchange.

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